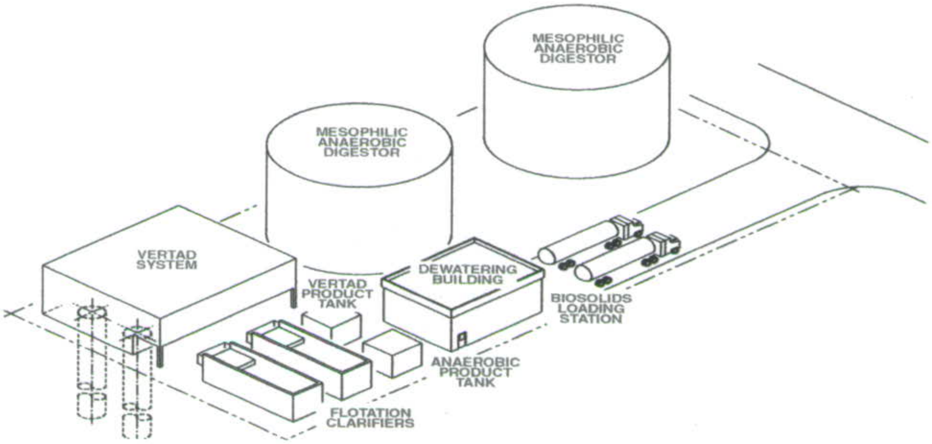
VERTAD™ Demonstration Project

AEROBIC THERMOPHILIC DIGESTION IN A DEEP VERTICAL REACTOR



FINAL
Evaluation Report

Prepared for:
King County Department of Natural Resources

Technology Supplier:
NORAM Engineering and Constructors Ltd.

Prepared by:
E&A Environmental Consultants, Inc.
19110 Bothell Way N.E. Suite 203
Bothell, WA 98011

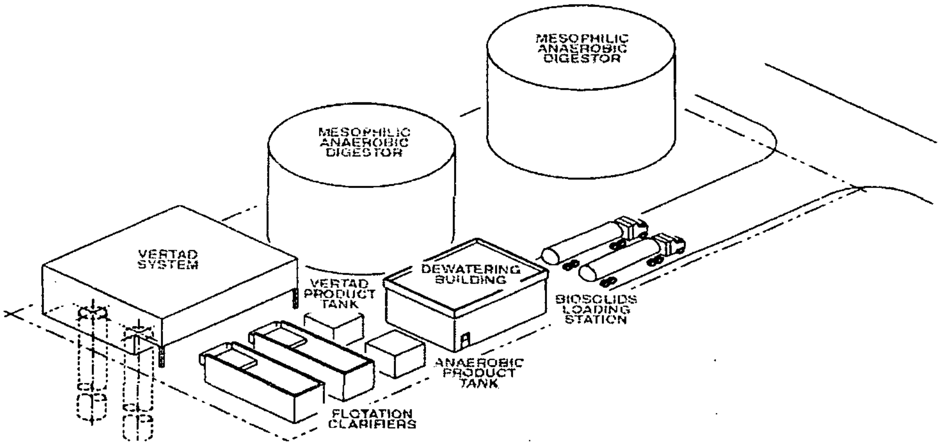
May 25, 2001

Project 30900

E & A Environmental Consultants, Inc.

VERTAD™ Demonstration Project

AEROBIC THERMOPHILIC DIGESTION IN A DEEP VERTICAL REACTOR



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March 28, 2001

Project 30900

Acknowledgment

This project began with submittal of a proposal to the Municipality of Metropolitan Seattle in August 1993. The project has been carried out by a team of process developers, designers, and operators with the invaluable participation and support of the King County Department of Natural Resources. The patience and dedication of those involved were essential for the successful completion of this technology demonstration project. Although all participants are not named here, the primary participants included the following:

King County Department of Natural Resources

Mike Boyle – Principal Investigator and Project Manager
Stan Hummel – Project Manager through the 1998 test period
Greg Bush – Program Manager
Rick Butler – South Plant Process Manager
John Smyth – Technology Assessment Program
Bob Bucher - Technology Assessment Program
Laboratory Support
Maintenance Crew Support
Operations Support

Deep Shaft Technology, Inc.

Dave Pollock, Process Developer and Project Engineer
Lyle Cuthbert, President

NORAM Engineering and Constructors, Ltd.

Dave Pollock, Process Developer and Project Engineer
Clive Brereton, Ph.D. - Chief Engineer
Jeff Guild – Primary Data Analyst
Steve Sapora - Vertad Cost Estimates

E&A Environmental Consultants, Inc.

Larry Sasser – Principal Report Author and E&A Project Manager
Chris Peot – Engineer and Troubleshooter
Joel Alpert – Quality Control

Staco Drilling, Inc.
Reactor Installation

Ohno Construction, Inc.
Civil Contractor

Harris Mechanical
Mechanical Contractor

Seven Sister Inc.
Electrical Contractor

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Executive Summary

Process Description

VERTAD™ is an aerobic, thermophilic, biological, digestion process performed in a 350-foot deep vertical reactor. The process produces a Class A biosolids product with desirable thickening and dewatering characteristics. The process requires electrical energy for aeration and produces hot water as a byproduct. It has been demonstrated at pilot scale to have beneficial synergistic effect when linked in series with mesophilic anaerobic digestion.

Primary Conclusions of Evaluation

1. VERTAD™ linked in series with mesophilic anaerobic digestion is the least cost alternative of the Class A solids processes considered in this study.
2. The digested biosolid product from VERTAD™ has a high potential for development of a local urban market, which would significantly reduce the cost of biosolids management. The product has a high fiber content compared to typical biosolids.
3. VERTAD™ has a similar net energy demand compared to Class A temperature-phased, anaerobic digestion.
4. The evaluation results indicate that the VERTAD™ technology merits the further testing needed to refine and expand the performance information that has been developed to date.

Primary Recommendations

1 It is recommended that the County conduct additional testing using the existing demonstration facility. This effort should focus on:

   a) Thickening the digested product using the Sulfuric Acid Flotation Thickening method.
   b) Dewatering with equipment that has demonstrated excellent performance with high fiber solids.
   c) Dewatered product quality for local urban use, including transformation by short term composting.
   d) Performance of linked VERTAD™ and mesophilic anaerobic processes.

2 The VERTAD™ technology merits consideration for full-scale implementation at both the South Treatment Plant and the proposed new North Treatment Plant.

Demonstration and Evaluation

of

VERTAD™
Aerobic Thermophilic Digestion Process

Section 1

Summary

The King County Department of Natural Resources has supported research and demonstration of wastewater treatment methods that have the potential to provide:

1. A smaller process foot print for solids handling facilities
2. Better control of odorous emissions
3. Reduced biosolids product hauling truck traffic

The objective of this investigation was to construct and operate a VERTAD™ reactor for the purpose of digesting mixed primary and secondary wastewater treatment solids using the aerobic thermophilic digestion process. The 350-foot deep VERTAD™ reactor has the ability to efficiently transfer large quantities of oxygen and to circulate and mix thickened wastewater solids. The process provides Class A Pathogen reduction and a stable biosolids product in a much smaller area than required by conventional digestion technologies. The study results include an economic analysis of the technology in comparison to other technologies being considered. This report provides an assessment of the development and performance of the VERTAD™ process through the initial investigation project. The findings of this evaluation are provided in the following summary.

VERTAD™ is a proprietary process for aerobically digesting wastewater solids with heat as a primary byproduct. The biologically generated heat allows the process to be operated at thermophilic (45-65°C) temperatures with no external heat source. Thermophilic temperatures speed the biological degradation process and provide compliance with the Class A biosolids criteria established by 40 CFR 503 and implemented in WAC 173-308.

Through sponsorship by the King County Department of Natural Resources, a demonstration facility was constructed, operated, upgraded, and operated again to demonstrate the performance of the VERTAD™ technology.

The facility consists of a 20-inch diameter by 350-foot deep reactor with a head tank that provides 5,535 gallons or 740 cubic feet of processing capacity. The facility can process 700 lbs of volatile solids per day or the equivalent to solids from a treatment plant processing 0.5 MGD of sewage. This would be typical loading from a community of 3,800 people. The demonstration unit required supplemental heat from a boiler to sustain thermophilic temperatures. This was due to the large surface area contributing to heat loss relative to the heat generating solids volume. In full-scale applications involving smaller diameter reactors such as this demonstration unit, the ground surrounding the reactor casing would be insulated, thereby preventing excessive heat losses. Excess heat would be generated from an uninsulated reactor with a four-foot diameter or greater.

The demonstration tests determined that the VERTAD™ process can produce a biosolids product that is Class A and dewaters to 30 percent total solids with a relatively low polymer dosage. These performance characteristics convert to significant cost reductions for dewatering and biosolids management relative to the current technologies and utilization program.

Concurrent tests completed by the University of Washington indicate that VERTAD™ product is digested very effectively by anaerobic digestion. This combination has the potential to optimize the performance of both processes to yield a lower cost processing method.

The economic analysis indicates that:

- The most cost-effective approach for using VERTAD™ is in combination with anaerobic digestion in order to produce a Class A biosolids product.
- VERTAD™ digestion alone is also a viable, although more costly, method of replacing anaerobic digester capacity at the West Point Treatment Plant and thereby reducing the digestion footprint.
- VERTAD™ is a cost-effective method of providing Class A biosolids processing at the South Treatment facility.
- VERTAD™, combined with a reduced capacity anaerobic digestion system is an economically feasible method of providing digestion and Class A product for the planned North Treatment facility.
- The process also has potential for use in processing source separated food waste as part of the County's organic waste management efforts.

Based on results of the process testing program and the economic analysis, it is recommended that continued evaluation of the VERTAD™ process be undertaken focusing on three critical aspects:

1. VERTAD™ product dewatering
2. Anaerobic and VERTAD™ process linking
3. The potential market for local usage of a Class A dewatered VERTAD™ biosolids product

1.1 Findings About VERTAD™ Economic Feasibility

A cost analysis was performed as part of this evaluation. The alternatives selected for comparison were based on the results of the demonstration testing program and on studies currently under way by the County to evaluate digestion processes suitable for use at each of the three regional treatment plants. Specific alternatives have been evaluated for each of the three facilities. Section 3 of this report provides a complete review of the cost effectiveness of using the VERTAD™ technology at each treatment plant. In general, VERTAD™ was found to be a viable digestion technology for use at both the South and North Treatment Plants. VERTAD™ could be used to reduce footprint at the West Point Treatment Plant, but at substantial cost.

The basic economic contrasts between the alternatives can best be summarized by using data for the North Treatment Plant. Since this is not an existing facility, the alternatives can be compared without cost effects from use of existing features.

The economic analysis considered five basic alternative digestion process configurations:

Anaerobic Processes:

Mesophilic (Anaerobic Base Case) – Traditional anaerobic digestion as currently practiced at the South and West Treatment Plants.

Thermophilic Class A – Anaerobic digestion in the thermophilic temperature range with a high temperature holding tank as required to produce a Class A product.

VERTAD™ Processes:

Concept 1 (NP1) – Provide full digestion (>40 percent solids reduction) in VERTAD™, then provide additional digestion in an anaerobic digester with reduced detention time.

Concept 2 (NP2) – Provide the minimum detention time needed to meet Class A requirements in VERTAD™, then complete digestion in anaerobic digesters.

Concept 3 (NP3) – Provide full digestion in a VERTAD™ reactor.

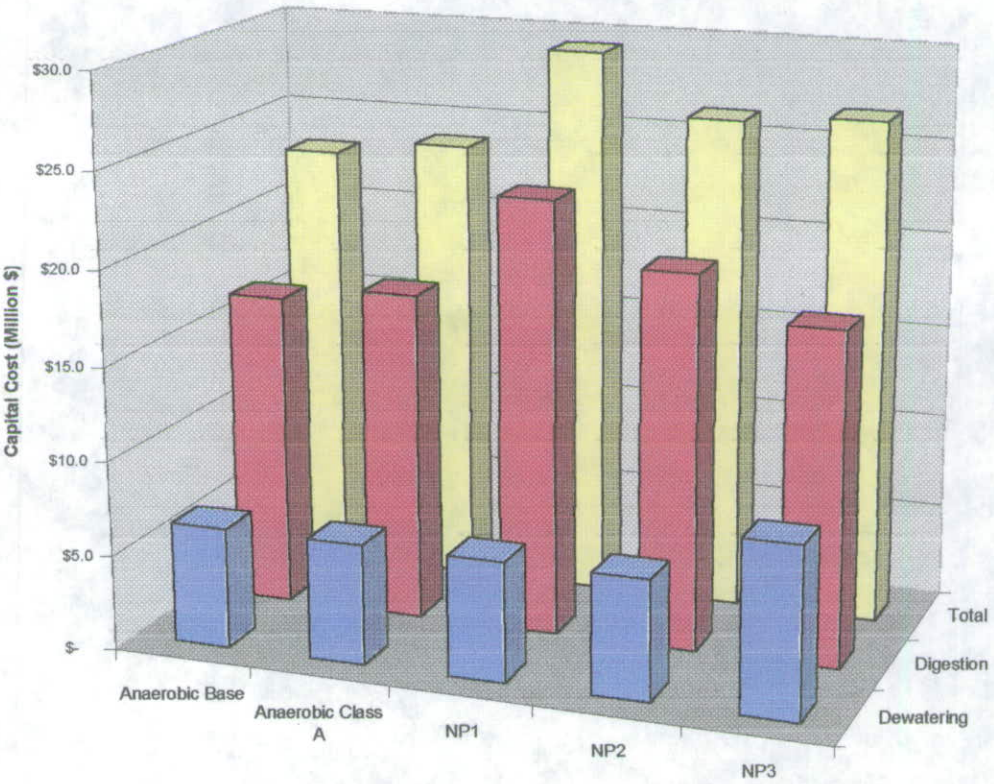
The cost analyses for the five alternatives include capital costs (digestion and centrifuge dewatering facilities), operating costs (energy, labor, chemicals, and biosolids distribution), and maintenance costs.

The costs for the anaerobic processes at the planned North Treatment Plant are based on estimates prepared by Brown and Caldwell for an additional digestion facility at the South Treatment Plant (South Treatment Plant - Enlargement III, Solids Treatment Enhancements - Alternatives and Recommendations, Task Report 740030, September, 2000). The VERTAD™ process alternatives were based on capital cost estimates provided by NORAM, and operation and maintenance costs for South Treatment Plant from the Brown and Caldwell report. The present worth of these costs was calculated for each alternative based on 17 years of operation to be consistent with the Brown and Caldwell report. An analysis of the sensitivity of the projected costs to changes in the assumed operating and capital costs was also conducted. The results are summarized in Graphics 1-10.

Using these five alternatives for the proposed North Treatment Plant, the differences between the processes become apparent.

Graphic 1 provides a comparison of the capital cost estimates for the five alternatives. The costs for the digestion and dewatering facilities associated with each alternatives are shown. All alternatives except NP3 (VERTAD™ digestion) have the same capital cost for dewatering facilities. NP3 produces more solids which results in higher cost. The lowest digestion costs are for the anaerobic processes and NP3. The combined anaerobic and VERTAD™ systems have higher capital costs.

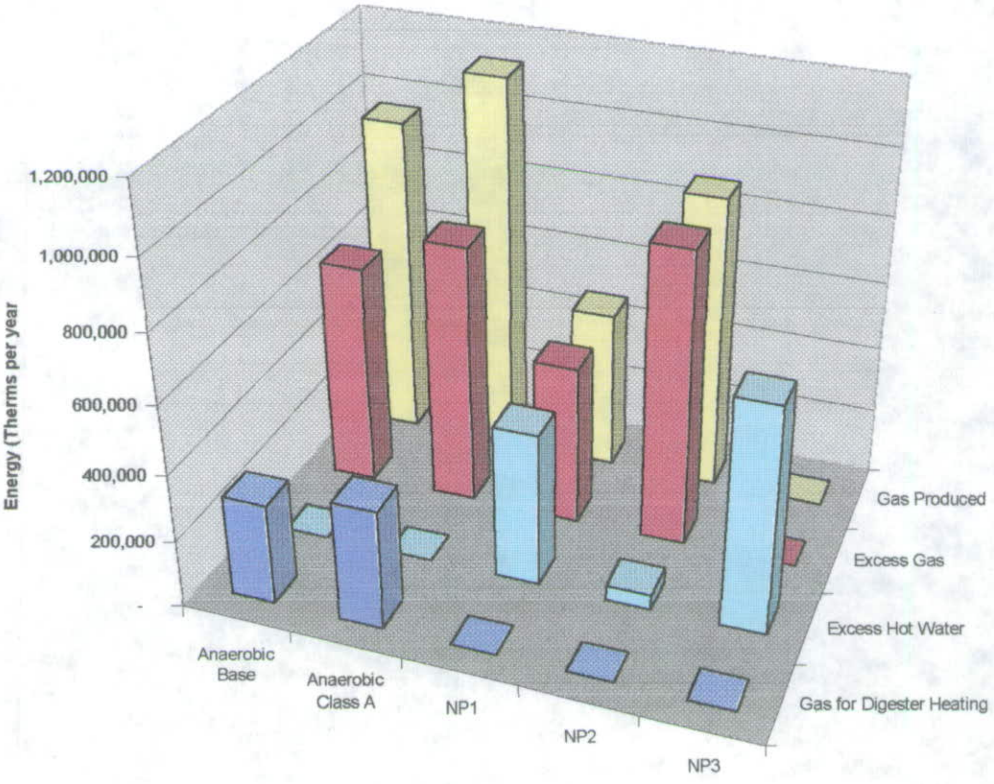
Graphic 1: Capital Cost Comparison



The results of the operation and maintenance cost comparisons give very different results than the capital cost comparison. The primary factors that resulted in lower operating costs for the VERTAD™ alternatives were related to energy value, polymer demand, and biosolids production.

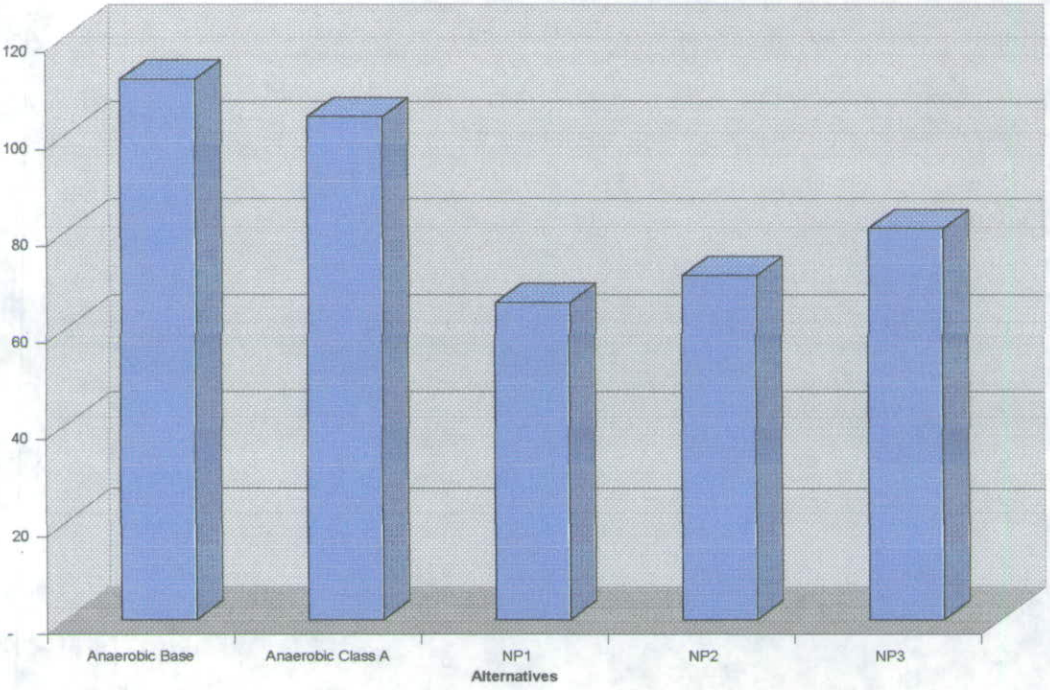
Graphic 2 shows, for each alternative, the anaerobic digester gas production, its demand for digester heating and the excess that is available for use elsewhere. The energy available in the form of hot water (a lower value fuel than gas) is also shown. Comparing gas energy produced by the alternatives with only anaerobic digestion (anaerobic base and Class A) against the alternative with VERTAD™ only (NP3) indicates that the anaerobic processes produce more energy and in a higher value form. However, a significant fraction of the energy produced in the anaerobic only alternatives is needed to heat the digesters. By combining the VERTAD™ and anaerobic processes in alternatives NP1 and NP2, the resulting available energy is increased above the anaerobic processes because the hot water produced by VERTAD™ is used to heat the anaerobic digesters. Similar amounts of excess energy are available from these two alternatives, but the excess gas energy produced by alternative NP2 has higher value.

Graphic 2: Gas and Hot Water Energy



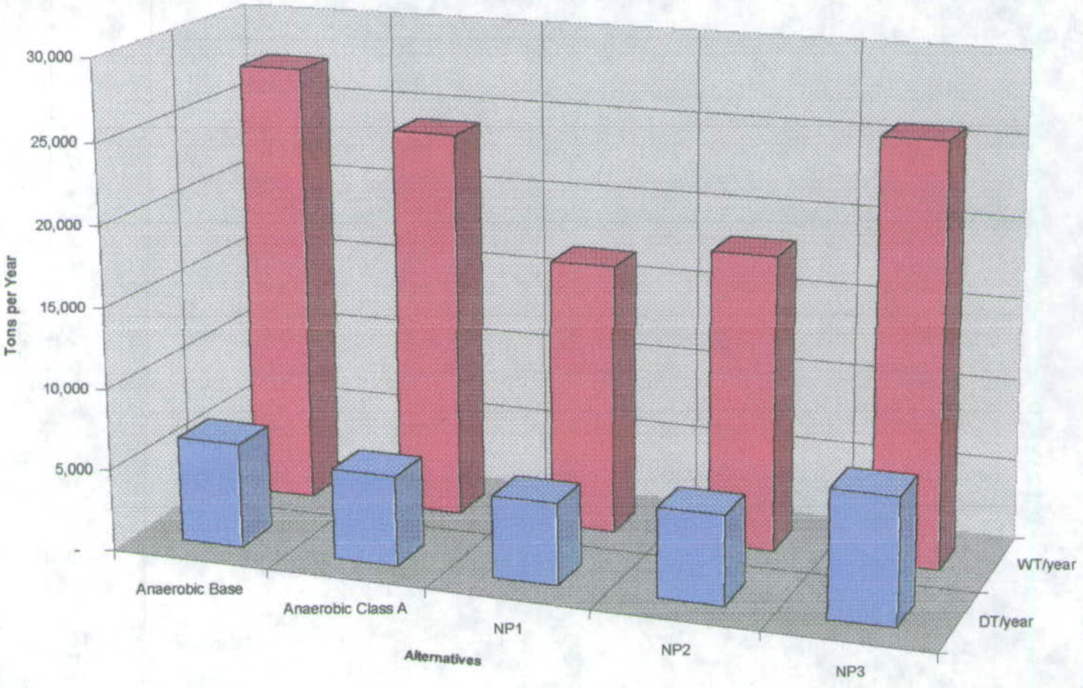
Dewatering is another primary differentiating factor. Graphic 3 shows the polymer demand for each of the alternatives. Dewatering performance tests on VERTAD™ product have indicated a significantly lower demand than experienced for anaerobic product. The polymer demand for the alternatives reflects a lower polymer demand and differences in volatile solids reduction between alternatives.

Graphic 3: Polymer demand (tons/year)



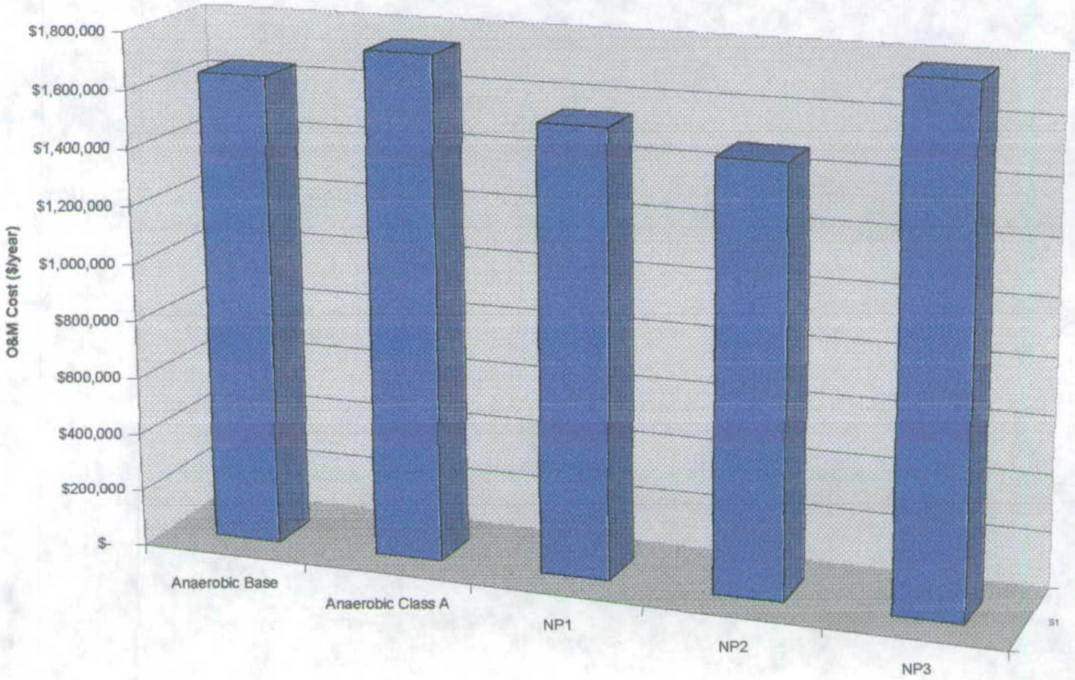
Biosolids production rates determine the cost for haul and application. Graphic 4 shows the effect of volatile solids reduction on the dry solids produced by each alternative. The quantity of dry solids is a direct reflection of the volatile solids reduction. The wet tonnage production shown on the same graphic reflects the dewatering effectiveness for the solids produced by each alternative.

Graphic 4: Biosolids Production (Tons per Year)



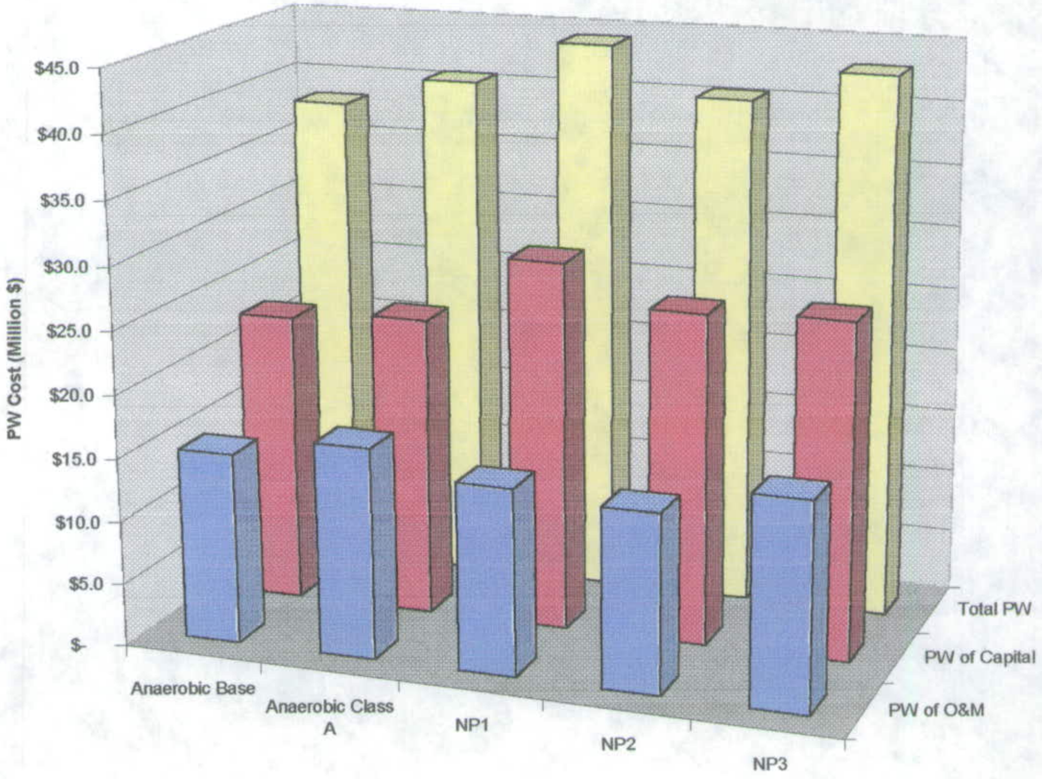
The effects of gas production, dewatering and haul and application together with all other factors provide the basis for the estimated total operation and maintenance costs presented on the following Graphic 5. The alternatives with the lowest operating costs combine the benefits of anaerobic and aerobic processes.

Graphic 5: Operation and Maintenance Cost Comparison ($/year)



Graphic 6 shows the present worth cost estimates for each of the alternatives. The present worth was calculated for the first 17 years of operation and added to the capital cost estimate. The present worth costs of capital are shown to be greater than the cost of operation and maintenance, and are therefore more influential in determining the alternative with the least present worth cost. The Class A anaerobic system and the VERTAD™ based alternative NP2 have essentially the same low total present worth cost, with NP2 having higher capital cost and lower operation and maintenance cost.

Graphic 6: Present Worth Cost Comparison (10^6$)

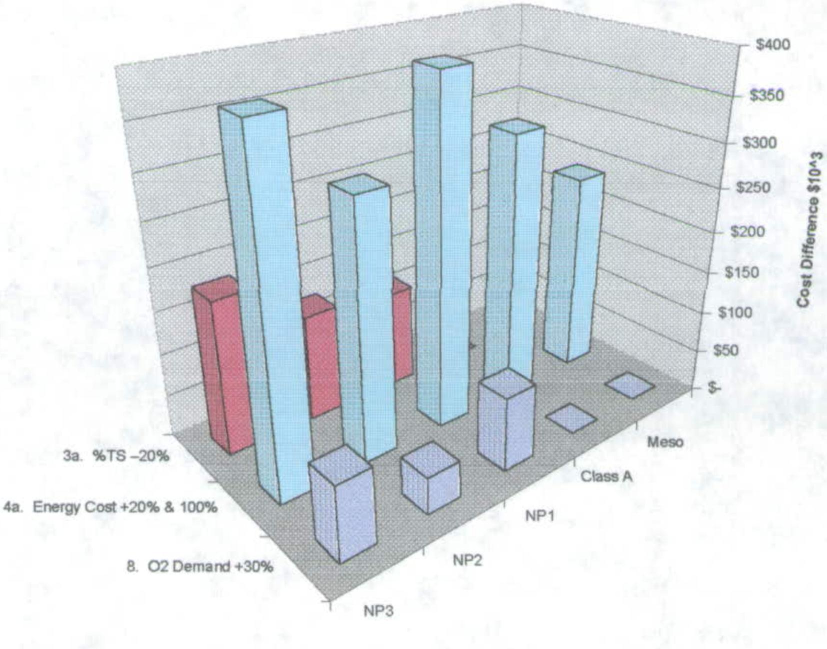


The anaerobic base case cost has the lowest present worth cost, but is the only alternative that does not provide a Class A product.

1.2 Cost Sensitivity Analysis Results

A cost sensitivity analysis was completed to evaluate the impact of changes in design criteria and costing assumptions on the cost estimates. Graphic 7 shows the cost sensitivity comparison results that produced the greatest increase in operation and maintenance costs. Reducing the cake *solids concentration assumption for the VERTAD™ alternatives from 30 percent to 24 percent* TS resulted in increased costs by from $110,000 to $180,000 per year. A 30 percent increase in the oxygen demand per volatile solids removed also increased the cost of VERTAD™ operations. An increase in the cost of energy also had a major impact on costs. The assumed increase was 100 percent for electricity cost and 20 percent for all other energy costs and values. Mesophilic anaerobic digestion, which processes a Class B biosolids product, was least impacted by increased energy costs. Among the Class A biosolids producing alternatives, the alternatives with full digestion in the VERTAD™ reactor had significantly higher net energy costs than Class A anaerobic digestion or the short contact VERTAD™ linked with anaerobic digestion.

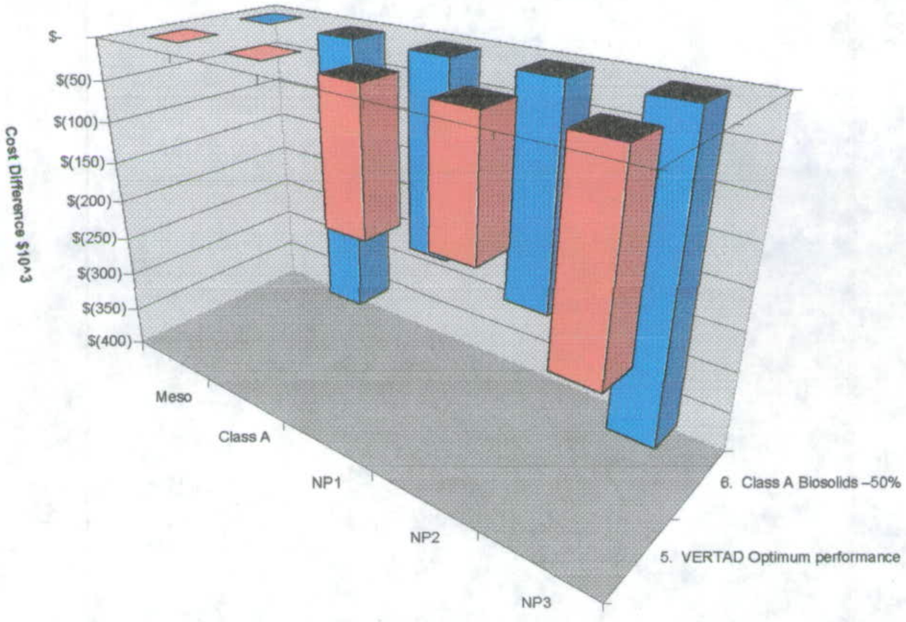
Graphic 7
Cost Sensitivity Analysis - O&M Costs
Alternatives with Increased Cost Impact



Graphic 8 shows the effect of two changes that significantly reduced costs. The largest impact resulted from assumed reduced haul and application cost associated with local use of Class A products. This is a potential cost savings and requires the availability of market capacity for Class A material.

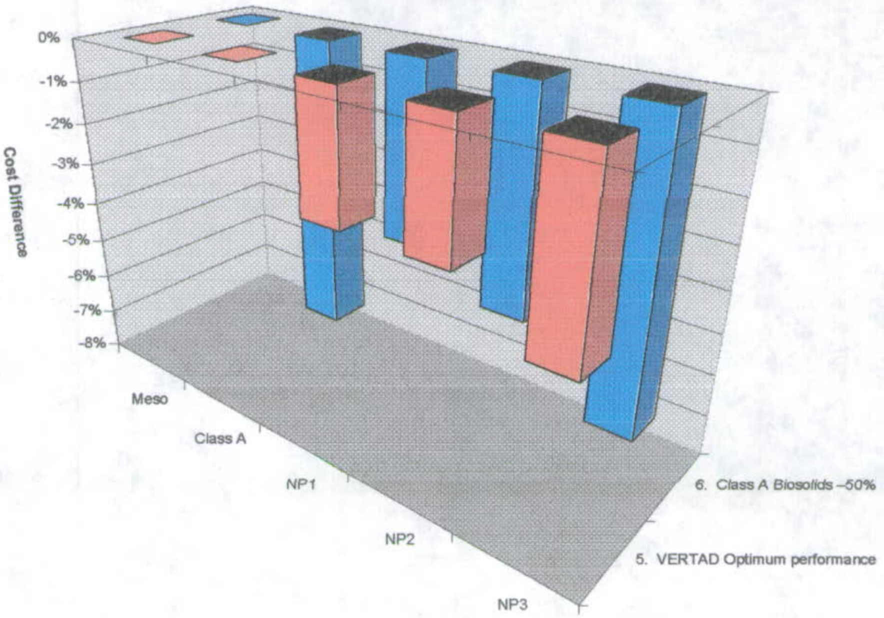
The market for Class A VERTAD™ product will need to be verified through additional analysis. The potential savings that may be available from optimizing the VERTAD™ process are also significant. Because of the level of development of the VERTAD™ process, relatively conservative assumptions have been made about performance. If performance levels demonstrated during testing can be shown to be reliably and consistently accomplished, then the cost of VERTAD™ operations will be significantly reduced.

Graphic 8
Cost Sensitivity Analysis - O&M Costs
Alternatives with Reduced Cost Impact



Graphic 10 shows the same effect on present worth that was indicated for operational costs. Both the production of Class A solids and operating at VERTAD™ optimum performance would have a significant impact on present worth costs.

Graphic 10
Cost Sensitivity Analysis - Present Worth Costs
Alternatives with Reduced Cost Impact



The sensitivity analysis indicates that several factors do not significantly impact the alternative cost comparison including operating labor costs, reduced volatile solids reduction in VERTAD™, and reduced gas production in VERTAD™ linked anaerobic systems. The analysis also shows that the primary focus for future consideration and evaluation of the VERTAD™ technology should be directed toward:

1. Dewatering
2. Markets for Class A product
3. Optimizing VERTAD™ performance
4. Refining costs for anaerobic digestion and VERTAD™ reactor placement

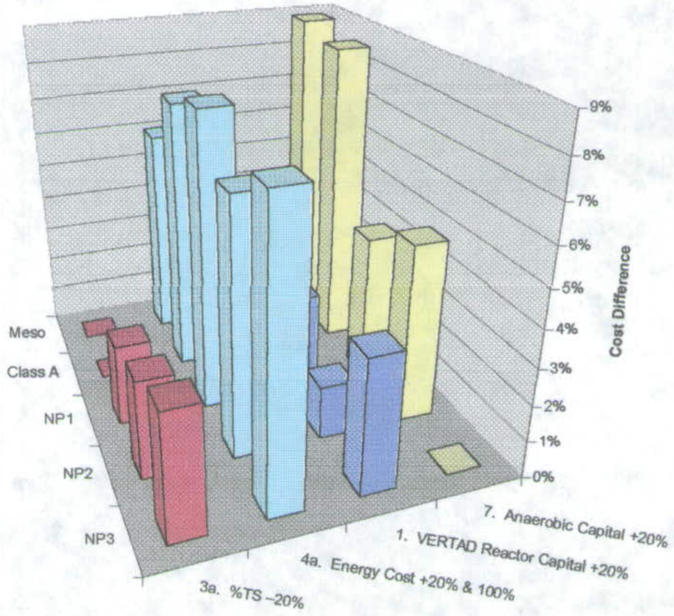
These factors should be investigated further prior to implementing any VERTAD™ based treatment plant improvements.

1.3 Performance And Economics Of The VERTAD™ Technology

VERTAD™ technology has been demonstrated to be a viable option for King County. Several attractive options for utilization of this technology by the County are available and appear to be technically feasible based on the information gathered to date. The process can perform as a stand-alone process, in sequence with anaerobic digestion (either mesophilic or thermophilic), or in combination with the Centridry process. In all cases the product will be Class A.

Graphic 9 shows the effect of changes that most significantly increase the present worth cost of some alternatives. The greatest impact results from a change in the capital cost of anaerobic digestion. A significant cost increase would also be experienced from an increase in the cost of installing a VERTAD™ reactor. Energy costs and VERTAD™ dewatering effects show the same impact previously discussed for operational costs.

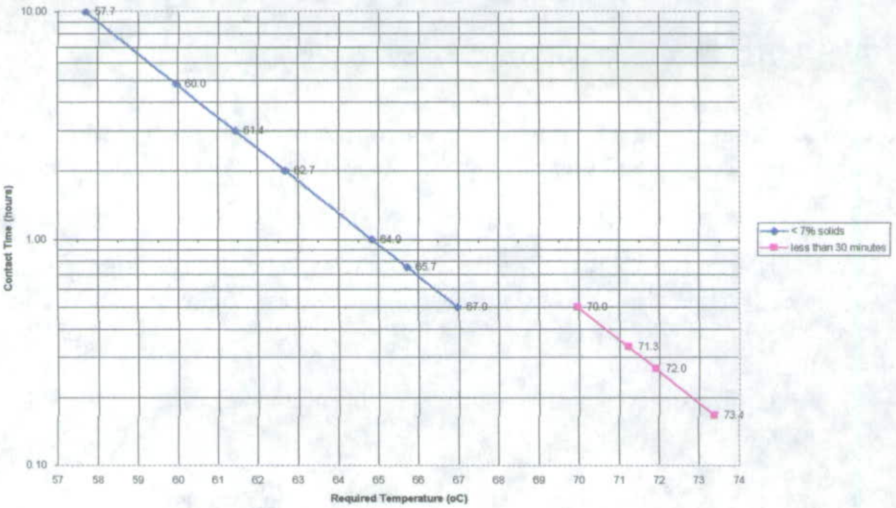
Graphic 9
Cost Sensitivity Analysis - Present Worth Costs
Alternatives with Increased Cost Impact



1.4 Findings About VERTAD™ Process Performance

The VERTAD™ process was demonstrated to produce a Class A biosolids product at a three to four day detention time. The following graphic shows the time and temperature requirement from the regulation. VERTAD™ would likely operate at about a 60°C temperature, which requires five hours of assured contact time.

Figure 1-1:
40 CFR 503 Class A Time and Temperature Requirements
for Solids Less Than 7%



The biological population operating in the conditions established by the VERTAD™ process has demonstrated reliability and resistance to upset during testing to date. Recovery from interruption of feed and supplemental heat has been consistently observed to occur within less than a day. The VERTAD™ process provided approximately 38 to 43 percent volatile solids reduction and approximately 35 to 48 percent chemical oxygen demand (COD) removal in a four day detention time at a 60°C average operating temperature. Oxygen transfer efficiency was measured at 30 percent to 62 percent during testing.

The process has the potential to greatly reduce the footprint required for digestion compared to anaerobic digestion.

Large scale (greater than four-foot diameter) VERTAD™ reactors operating at a four day detention time on thickened feed will generate excess heat recoverable as hot water that can be used for space heating and for digester heating in linked anaerobic systems.

Bench scale testing indicates that the VERTAD™ product dewaters significantly better than anaerobic product, and if pre-float thickened, requires less polymer.

1.5 Findings About Test Facility Performance

Operations during the first two test series indicated the need for more sophisticated control and monitoring capabilities than was originally anticipated. Improvements were made to level sensors and controls in the process tanks and the data collection system was automated. These improvements allowed demonstration of the potential usefulness of the technology for King County during the third test series. Additional improvements before the last test series provided more reliable heat addition, improved mechanical and electrical reliability and remote monitoring and data collection capabilities. Improvements to the demonstration facility have improved the reliability of the facility and the performance data generated. However, despite the improvements made to date additional changes will be needed before additional testing should proceed.

1.6 Findings About Additional Testing Needs

The combined results of the performance testing and cost estimates that have been developed during this evaluation indicate that any further evaluation of the technology should focus on three primary areas:

1. Dewatering of the VERTAD™ product for the following configurations:
   a) VERTAD™ followed by sulfuric acid flotation thickening (SAFT)
   b) VERTAD™ followed by SAFT followed by anaerobic digestion (mesophilic and thermophilic)
   c) Anaerobic digestion (mesophilic and thermophilic) followed by VERTAD™ followed by SAFT

2. Linked anaerobic and VERTAD™ process configurations

3. Class A product market potential should be evaluated to determine how much Class A product could be used locally. Development of a local market for Class A product would dramatically reduce hauling and utilization costs.

Demonstration and Evaluation

of

VERTAD™

Aerobic Thermophilic Digestion Process

Section 2

Technology and Demonstration Development and Demonstration Facility Performance

This section of the report provides background information of the VERTAD™ process and development of the demonstration facility. The discussion then proceeds to describe the demonstration facility, the testing plan and the results of the facility operations performance.

2.1 Technology Development History

The VERTAD™ process has evolved from the Vertically Oriented Aerated Pressure Cycle Bioreactor developed by ICI in Great Britain. The technology was initially developed for the aerobic fermentation of methanol for single cell protein concentration. Researchers recognized potential benefits from using the technology for wastewater treatment. The deep reactor technology was first used for wastewater treatment as an activated sludge process in 1974. The North American rights to the technology were licensed to CIL, Inc. in 1975. CIL continued development of the technology. Full-scale facilities using the technology for activated sludge were constructed in Great Britain, Canada and Japan. The technology was sold to Simmons Group of Calgary in 1986. Simmons participated in the installation of the first full-scale facility in the United States at Homer, Alaska in 1990. Simmons developed the VerTreat (activated sludge) and VERTAD (aerobic digestion) technologies during the early 1990's. A proposal to test the technology was submitted to the Municipality of Metropolitan Seattle in 1993, which resulted in the project that is the subject of this report. A full-scale facility using both technologies (activated sludge and mesophilic aerobic digestion) was recently completed for use by Chevron in Burnaby BC. Rights to the VERTAD™ technology was acquired from Simmons by NORAM Engineering and Constructors Ltd. in September 1998.

The proposal to implement a demonstration project of the VERTAD™ aerobic thermophilic digestion was accepted in 1994. Prior to project initiation the County conducted an assessment of the potential for earthquake damage to a deep reactor. The study concluded that damage to the reactor would likely be less than to surface tankage (Ref. 3). No potential negative impacts on groundwater were identified. Construction of the demonstration facility was begun with drilling and installation of the reactor in October of 1996. Construction of the facility was completed in December of 1997. Start-up began on January 15, 1998 with operating conditions first achieved on January 21, 1998. Facility refinements delayed the first intensive test period until April 1998. The first operating period consisting of two tests was completed May 6, 1998. Additional facility changes were made prior to the third operational test, which was completed on December 17, 1998. Additional changes were completed in late July 1999. The latest series of operational tests were completed on Dec. 20, 1999.

The VERTAD™ technology is designed to be a simple, reliable and cost effective method for providing stabilization of thickened wastewater solids. The VERTAD™ demonstration project is supported by King County's Technology Assessment Program (TAP) (formerly called Applied Wastewater Technologies Program (AWT)). TAP is a research program with the objective of evaluating and testing technologies to reduce the environmental impacts of treatment plant operations. Specific objectives of the TAP program include reducing:

1. The space required by solids handling.
2. Biosolids truck traffic.
3. Odor emissions.

The project team led by E&A Environmental Consultants, Inc. (E&A) was responsible for the planning, design, construction, and initial operational testing of the facility. The technology owner, NORAM Engineers and Constructors Ltd. (NORAM), is actively involved in all aspects of the testing program. The County has provided engineering, operations, and maintenance support throughout the project. Following initial testing, the facility was turned over to the County. All subsequent modifications to the facility have been completed by the County.

The primary objectives of the initial testing program were to evaluate:

1. The VERTAD™ reactor with regard to hydraulics, oxygen transfer, and energy balance.
2. The capability of the process to stabilize the solids, provide compliance with the Vector Attraction Reduction and Class A pathogen requirements of 40 CFR 503 and Washington Administrative Code 173-308.
3. The process loading and detention times that will provide dewatered solids quantities comparable with current County facilities.

The anticipated digester detention time of two to eight days would represent a significant reduction in processing volume compared to the anaerobic mesophilic digesters currently used at King County treatment plants. The reduced detention time together with the vertical subsurface construction has the potential to dramatically reduce the space required for solids stabilization.

This space conserving technology for production of Class A biosolids uses basic biological treatment concepts that are already familiar to wastewater treatment operators. The test facility is actually a small full-scale test facility consisting of a 20-inch diameter by 350 foot deep sealed reactor tube (capacity estimated at 500 to 2,000 lbs. of solids per day) together with support appurtenances.

This technology is believed to have the following benefits:

- Provides a Class A biosolid product (40 CFR 503.32 Alternative 1)
- Vector attraction reduction (VAR) by reducing the volatile solid content by 38 percent
- Flotation thickening using dissolved gases in the product
- Dewatering to high solids content with low polymer demand following flotation thickening.
- Minimal footprint requirement
- Highly efficient (low energy demand) oxygen transfer
- Enhanced microbiological degradation due to efficient, high energy mixing

- Construction using conventional well drilling equipment or common mining techniques for large diameter reactors.
- Low volumes of process air delivered by efficient compressors
- Simple open pipe aeration device requires no maintenance
- Contained low volume off gas that is treatable for odor reduction

2.2 Demonstration Facility Construction and Improvements

The facility is located at the South Treatment Plant (STP) operated by King County in Renton, WA. The STP is a 125 MGD facility with primary clarification, activated sludge, dissolved air flotation thickening, anaerobic digestion and belt press dewatering. A summary of the design features for the VERTAD™ facility is provided on Table 2-1. The key component of the facility is a reactor constructed in a 350-foot deep, 20-inch diameter vertical tube. Some features of the test facility have been modified during the testing period. This description covers the initial test unit. Changes will be described. The reactor tube was placed by conventional drilling technology using the dual air rotary drilling method. Subsurface geology found during drilling consisted of 110 feet of coarse sand and gravel alluvium above a bedrock of siltstone and shale for the balance of the depth. There were indications of flowing water above the bedrock. Groundwater levels in the area are known to be 10 to 18 feet below the surface.

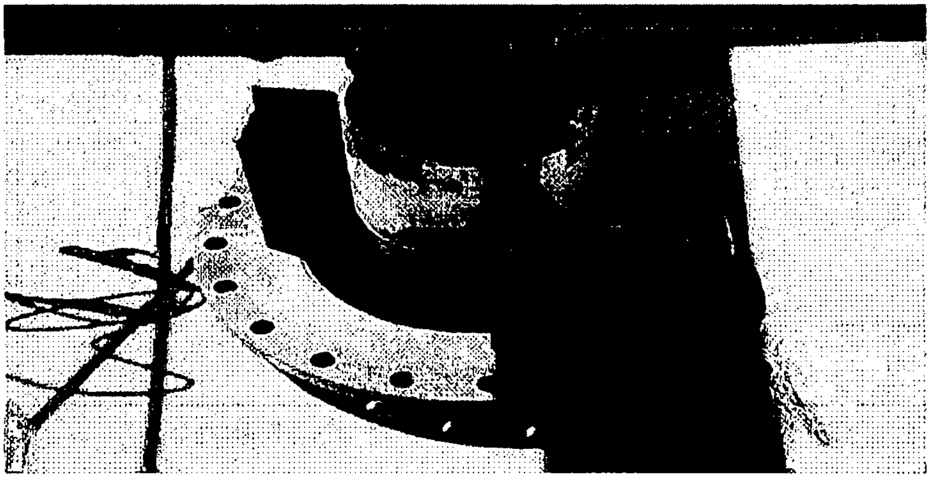
Table 2-1: VERTAD™ Demonstration Project Facility Design Summary

Influent Characteristics	*Value*
Influent	Thickened municipal biosolids
Loading	500 to 1,500 lb. solids/day
Solids Concentration	6.5 %
VSS	78 - 80%
Primary Sludge	60%
WAS	40%
Temperature	20 - 21°C
Biofilter treatment - building exhaust	800 cfm
Biofilter loading rate	5 cfm/sf
Feed Rate @ 3 day HRT	1,770 gallons/day
@ 6 day HRT	889 gallons/day
Equipment Inventory	
VERTAD™ BIO-REACTOR	
Casing	1 @ 20 in. dia by 350 feet deep
Draft tube	1 @ 10 in. dia by 143 feet deep
Extraction	1 @ 3 in. dia by 347.5 feet deep
VESSELS	
Feed Tank	1 @ 60 in. dia by 72 in high
Digester Head Tank	1 @ 60 in. dia by 72 in high
Purge Water Tank	1 @ 38 in. dia by 48 in. high
MECHANICAL	
Aeration Compressor	87 scfm, 150 psi, 25 HP
Feed Pump	1-10 gpm, 50psi, 3 HP
Purge Water Pump	20 gpm, 26 foot TDH, 5 HP
Reactor Volume Total	740 cf
Voidage	30 cf
Liquid	710 cf

The reactor surface feature is a five foot diameter tank (head tank) operating at a pressure of 3 to 5 psi. Figure 2-1 is a process and instrumentation drawing for the test facility. Figure 2-2 shows the floor plan.

The reactor is divided into sections in which thickened wastewater solids being processed are first circulated in an upper complete mix zone (surface to 160 feet depth). Flow in this zone is induced with an air-lift pumping action by injecting air into the reactor and circulating down through a 10 inch diameter tube and up through the annular space. Photo 1 shows the top of the downcomer and annular space during construction.

Photo 1: Downcomer Supported on Reactor Flange



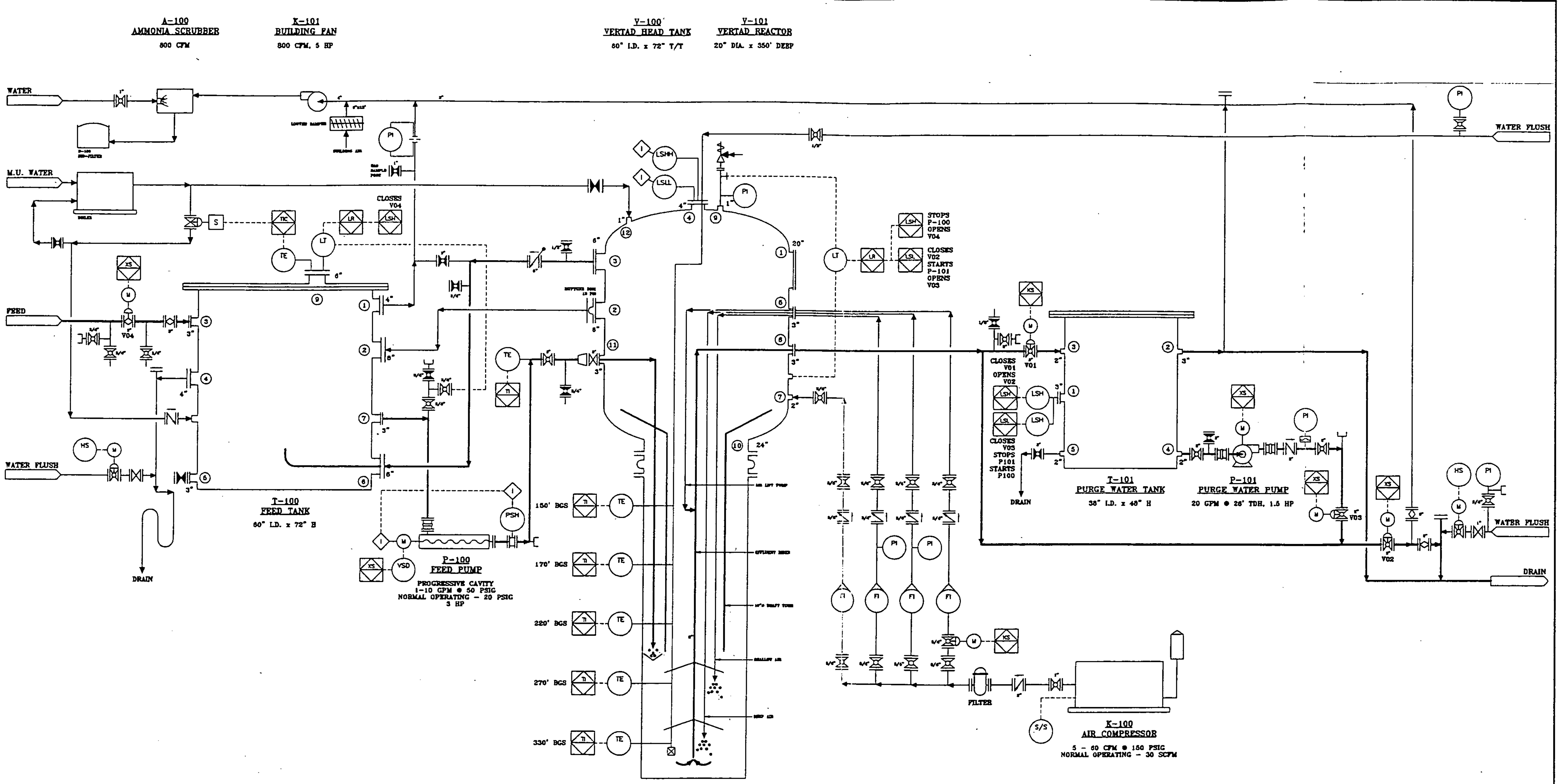


Figure 2-1

REV.	DESCRIPTION	DATE
A	REDRAWN AS EXISTING	FEB11.99

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VERTAD DEMO PROJECT
RENTON, WASHINGTON

PIPING AND
INSTRUMENTATION DIAGRAM

NORAM Engineering and Constructors Ltd.

SCALE: NTS	APPROVED BY:	DRAWN BY: PR
DATE: FEB1/99	CHECKED BY:	REVISED:
DRAWING NO. B-312-1401	SHEET 1 OF 1	REV. A

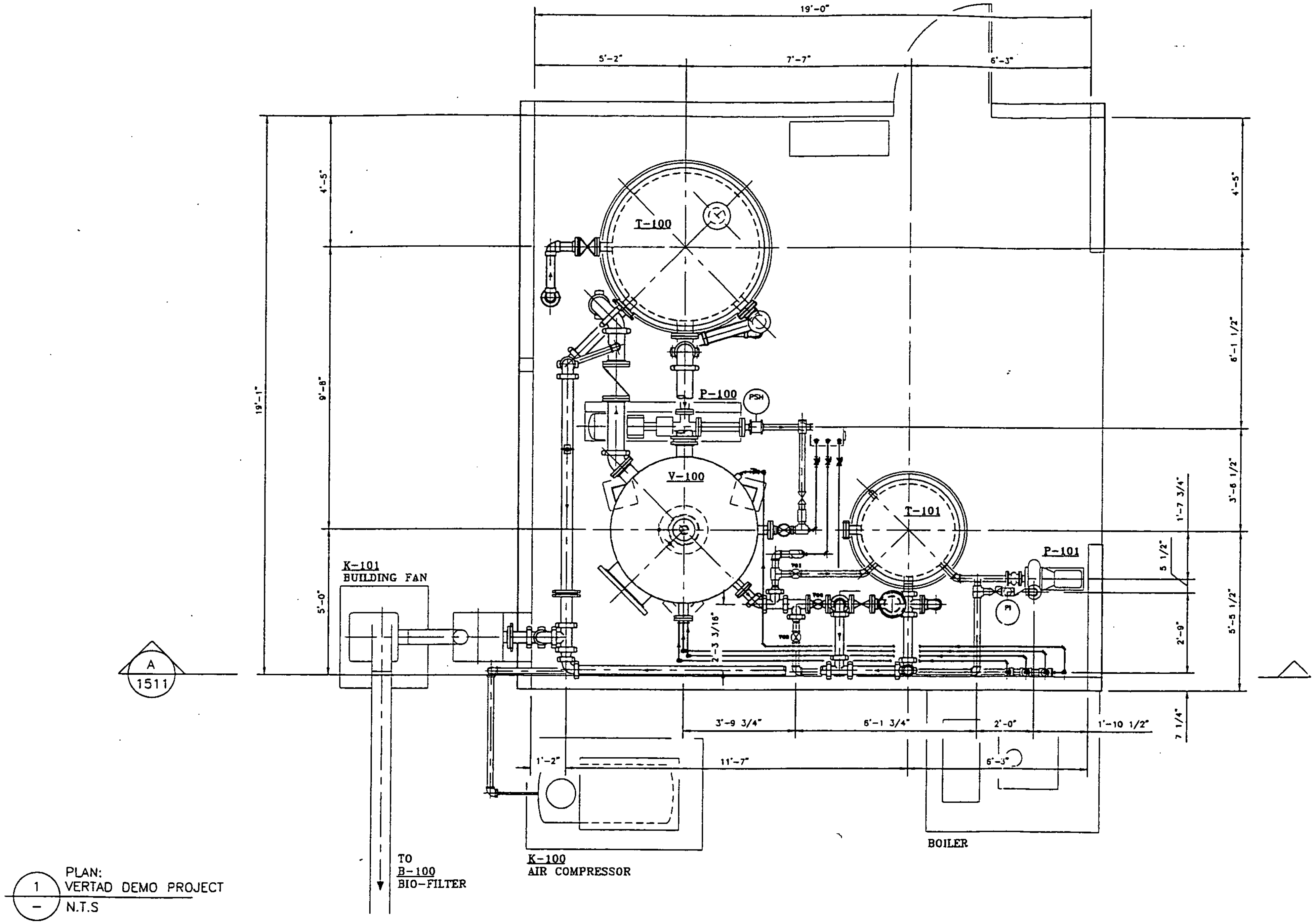

Figure 2-2

REV.	DESCRIPTION	DATE
A	REDRAWN AS EXISTING	FEB11.99

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VERTAD DEMO PROJECT
RENTON, WASHINGTON

PIPING & EQUIPMENT
LAYOUT

NORAM Engineering and Constructors Ltd.

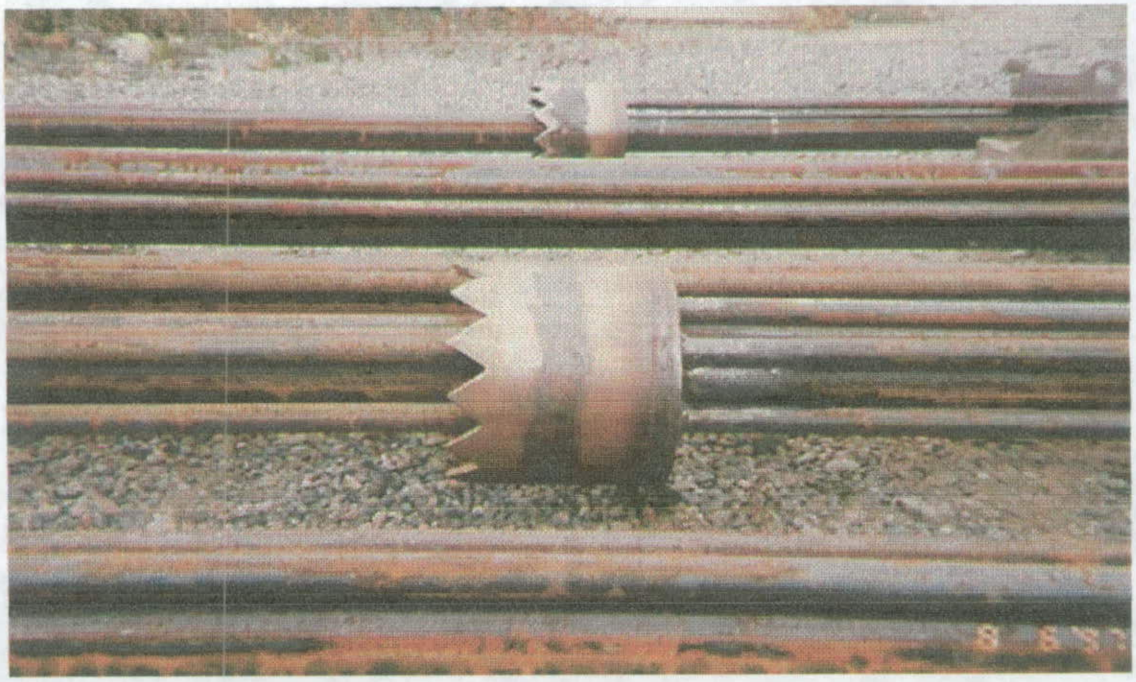
SCALE: NTS	APPROVED BY:	DRAWN BY: PR
DATE: FEB09.99	CHECKED BY:	REVISED:
DRAWING NO. B-312-1510	SHEET 1 OF 1	REV. A

The second section is designed for plug flow while being aerated from the bottom of the zone (160 to 315 foot depth). The mix is saturated with oxygen in this zone which experiences pressures of 5 to 10 atmospheres. The final "soak" zone (315 to 350 foot depth) is designed for plug flow without aeration. The biomass utilizes the residual oxygen from the lower aerated zone as it passes through this final degradation and pathogen control zone. The reactor is completely sealed to prevent impacts on groundwater. Photo 2 shows the building that houses the facility and the biofilter. The rotary screw compressor is at the left side of the building. The boiler had not been added at the time this picture was taken.

Photo 2: VERTAD™ Facility

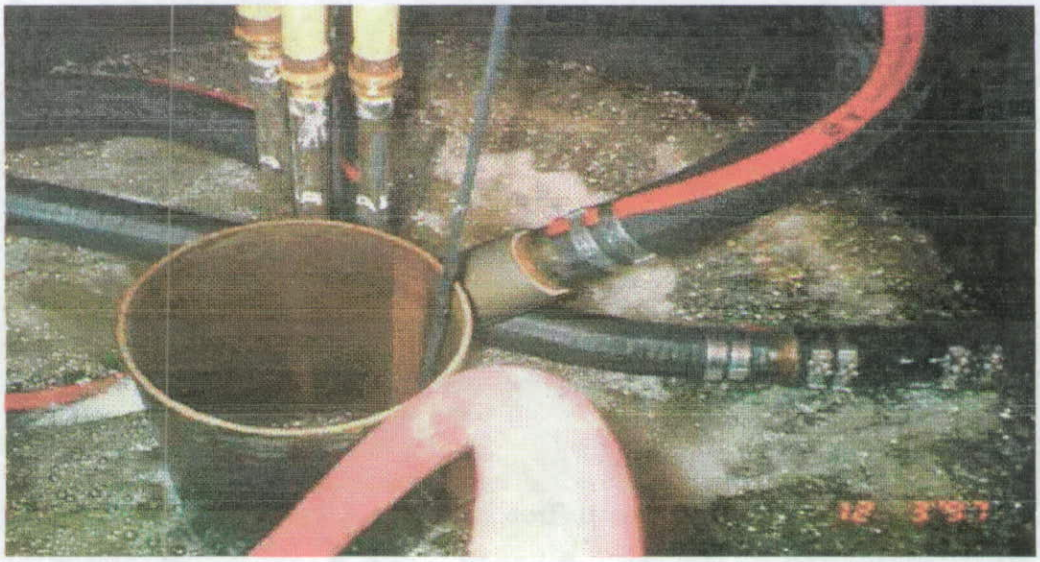
The support equipment for the reactor includes a thickened solids (THS) supply loop, feed tank, feed pump with variable frequency drive, purge water system, compressor, steam boiler, programmable logic controller (PLC), and biofilter. The batch product withdrawal and feeding cycles are fully controlled by the PLC. The THS supply loop provides a continuous supply of fresh solids (60 percent primary and 40 percent secondary) from the ESRP solids system storage tank. The 78 cf feed tank provides back pressure to the head tank, captures foam from the process and accepts steam for supplemental heating of the feed stream. THS enters the feed tank on a batch basis controlled by float switch level sensors. The 10 gpm variable flow feed pump batches feed to the reactor on a programmed schedule. The quantity fed with each batch is set with differential pressure level sensors in the head tank. Product is withdrawn from the bottom of the reactor through a three inch diameter extraction line. To prevent clogging between batches, purge water can be held in the extraction line between withdrawals. In this mode, the purge water is transferred to a 17 cf purge tank. Following product discharge the purge water is pumped back into the extraction line. Process air is injected at two elevations. The upper and lower aeration heads are 158 feet and 319 feet below floor level respectively. The aeration headers and extraction line assembly prior to installation is shown on Photo 3.

Photo 3: Extraction and Aeration Line Assemblies with Aeration Headers



The aeration headers supply the air needed to supply oxygen for the reaction and to mix the contents of the reactor. Air that is not dissolved results in bubbles which create voidage (volume of bubbles per unit volume of water). Photo 4 shows the aeration bubbles rising at the head tank periphery and flow moving toward the downcomer.. The float switch stilling well is in the center left, and the extraction line and feed line are on the right. The air lines are just above the stilling well

Photo 4: Reactor Internals with Aeration and Circulation



Aeration is provided by a continuous duty, rotary screw compressor. The unit provided requires 25 HP to produce 87 scfm @ 150 psi. The initial steam boiler provided a maximum of 80,000 BTU per hour of steam heat to supplement the process's biological heat production. The boiler was provided rather than equipment to recover heat from the product. Sensors provide continuous recording of temperatures at 150, 170, 220, 270, and 330-foot depths in the reactor. The test facility is housed in a temporary building that is provided with utilities and building air and off-gas collection and processing through a biofilter.

Because voidage in the reactor liquid column creates lower fluid density in the column than the liquid in the extraction line (which does not contain bubbles and is denser), special design consideration is needed to induce product flow through the extraction line. To force flow through the extraction line, the pressure in the annular space must exceed the hydrostatic pressure in the extraction line plus any head losses due to flow. Flow in the extraction line is induced by using a weighted check valve to create an additional 1-5 psi of head tank pressure. This allows the upper and lower zone voidage head to be 12-24 feet. Normally the lower zone contains relatively little voidage (6-12 ft.) because the bubbles are under 5-10 atmospheres of pressure. The upper zone voidage has been reduced by circulating large amounts of degassed head tank liquor through a 158 ft. long recycle pipe suspended in the center of the upper reactor. The upper zone voidage, despite the much larger volume of air due to the lower upper region pressure, is also 6-12 feet.

Extraction line flow can also be created by injecting air into the extraction line through the air lift injection port provided at the 240 ft. level. Using airlift can reduce or eliminate the need for head tank back pressure. The use of airlift, however, prevents monitoring of dissolved oxygen of the effluent stream and voidage changes (circulation changes) within the reactor.

Excessive heat loss has been a continuing problem for the process. The reactor was not insulated and has a high surface area to volume ratio, which facilitates heat transfer to the environment. Flowing water was identified in three zones during drilling. Water moving past the reactor can remove substantial heat. To compensate for heat loss to the environment, reactor feed was initially preheated via steam injection using an 80,000 btu/hr. propane-fired steam boiler. After the 1999 improvements a hot water propane boiler and heat exchanger loops installed in the reactor shaft were used to supply supplemental heat.

Experience from the initial test periods indicated the need for improvements to the process control and monitoring equipment. In addition to the new boiler, improvements installed in 1999 included:

- Installation of differential pressure level sensors in the head and feed tanks
- Direct monitoring and logging of the temperature, steam valve status and levels in the feed and head tanks through the PLC to a PC
- Direct PC oversight of the PLC control program
- Remote access to the PC for monitoring of process status
- Installation of electrical feedback protection for the PLC
- Analysis for additional operating parameters
- Addition of gas analysis capability
- Installation of temperature sensors with PLC monitoring of the feed tank, feed line, and in the reactor at depths of 150, 170, 220, 270, and 330 feet below ground surface

2.3 Demonstration Facility Operating History and Performance Information

The facility operations history to date is summarized as follows:

- Construction Completion - December 15, 1997
- Hot water heating period - December 18, 1997 to January 14, 1998
- Biological start up without heat - January 15 to January 21, 1998
- Temporary heat source operation - January 22 to February 4, 1998.
- Autothermal conditions (minimal supplemental heat) - January 29 to February 4, 1998.
- New boiler operational, February 6, 1998.
- Operating to develop stable process, February 6 to April 5, 1998.
- First intensive monitoring period three-day HRT, April 6 to 8, 1998.
- Transition and stabilization, April 11 to May 3, 1998.
- Second intensive monitoring period, May 4 to May 6, 1998.
- Shut down for feed pump failure, May 7, 1998.
- Mechanical, control, and monitoring upgrade, May-Oct. 1998
- Clean water testing, November 4 to 9, 1998.
- Biological start up, November 10, 1998.
- Process stabilization, November 11 to December 12, 1998.
- Third intensive monitoring period, December 13 to 17, 1998.
- Completion of initial tests, December 17, 1998.

- Completion of hot water boiler and heat exchange tubing installation, July, 1999
- Completion of control and data generation and capture capabilities, July, 1999
- Final operating test period, August through December, 1999

2.4 Demonstration Test Plan

The testing program was designed to determine the performance of the VERTAD™ process relative to the objectives of the King County TAP program. This includes identifying design criteria suitable for use in sizing and evaluating a full-scale treatment system. The particular parameters of interest include:

1. Fecal coliform or Salmonella to document compliance with WAC 173-308-170 and 40CFR503 requirements for Class B pathogen levels needed for land application of biosolids product. Exposure time and temperature to document compliance with Class A requirements for unrestricted use of the biosolids would be a highly desirable side benefit.
2. Volatile solids reduction to document compliance with WAC 173-308-180.
3. Operating hydraulic detention time and temperature at which reliable and predictable performance is provided by the process.
4. Rate of air delivery and delivery pressure to sustain the process.
5. Product dewaterabililty, polymer requirement, and sidestream quality.
6. Suitability of product for land application (stability, appearance, odor, and nutrient regulated metal content)
7. Odor control by biofiltration.

In addition, the testing program is designed to provide a more in-depth understanding of the process that may lead to future improvements in process performance and reliability.

The approach used to document process performance includes a reduced level of routine monitoring during the process stabilization period. The process is operated at a constant loading condition to demonstrate stable operation. Once stable operation has been accomplished for three hydraulic detention times (or solids retention with recycle), an intensive sampling and analysis period of four days is used to define the performance of the process. The monitoring program varied somewhat between tests. A representative monitoring program for VERTAD™ testing is provided as Table 2-2.

Table 2-2: VERTAD™ Mass Balance Monitoring Schedule

	Sampling Frequency - Test Period									
	Flow	*TS*	*VS*	*NH4*	*Alkalinity*	*TKN*	*COD*	*TOC*	*TC*	*FOG*	*Dewater*
THS	Daily	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	1/week	
Steam	Daily										
Condensate	Daily										
Feed Tank Overflow	Daily										
Feed Tank Off Gas	Daily										
Reactor Feed	Daily	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	1/week	
Air	Daily										
Head Tank	Daily	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c		
Product	Daily	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	Daily-c	3/week	1/analysis
Reactor Off Gas	Daily										

c-designation indicates the analysis of a composite of 3 grab samples
pH, ORP, DO, and gas analysis each batch minimum with intensive through one cycle per day.

During the intensive monitoring period, samples were collected for laboratory analysis of the thickened solids (THS), feed tank and head tank (upper zone) materials, and final product. These samples were analyzed for total and volatile solids (TS and VS), total Kjeldahl nitrogen (TKN), ammonia (NH_4), nitrate, alkalinity, pH, and ortho-phosphate by the ESRP laboratory. During the 3-day duration intensive monitoring period, samples were collected for each discharge cycle to evaluate sample variability. The King County Environmental Laboratory also analyzed daily samples for fecal coliform and Salmonella. During the third test series additional laboratory and field testing was conducted including chemical oxygen demand (COD), fats, oils, and grease (FOG), total carbon (TC), total organic carbon (TOC), off gas analysis, density, oxidation reduction potential (ORP), dewaterability, and dissolved oxygen. Product samples were also sent to dewatering equipment suppliers for dewaterability testing.

2.5 Demonstration Testing Performance

VERTAD™ process performance characteristics were evaluated during three phases of testing:

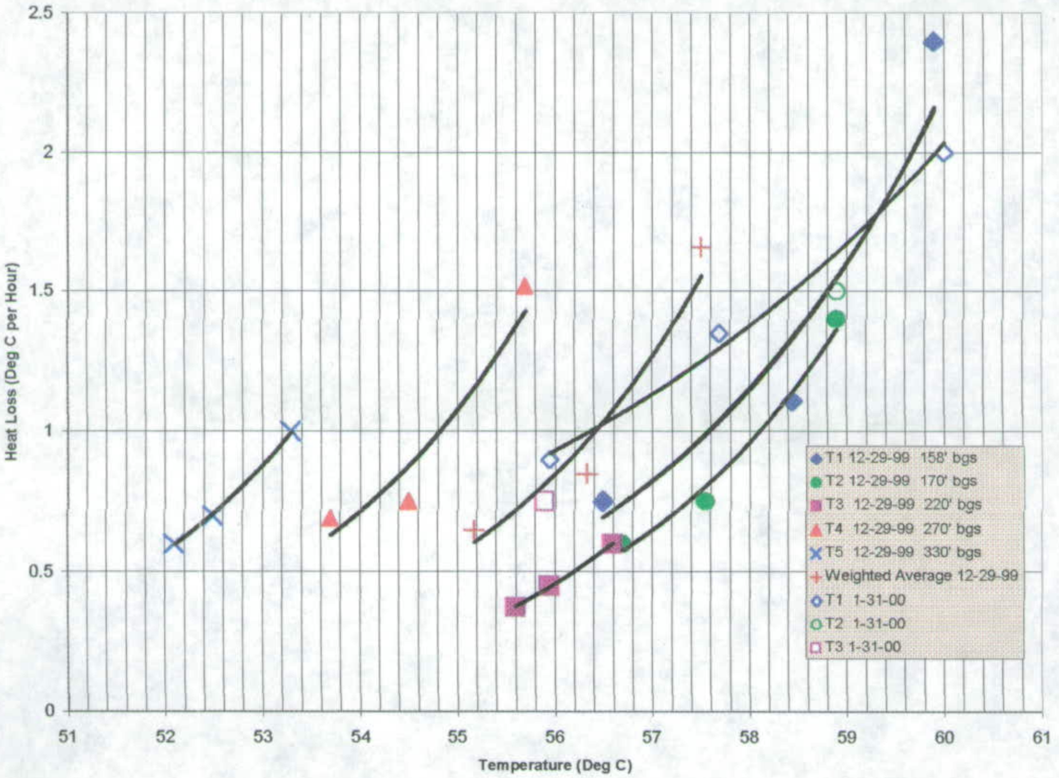
1. Clean water operation and substrata heating.
2. Biological process start-up and initiation of thermophilic activity.
3. Operations at predetermined operating conditions.

Each of these phases provide important information and insight into the performance of the VERTAD™ process.

2.5.1 Clean Water Testing

Prior to the addition of solids at the beginning of each test phase clean water tests were conducted to document reactor mixing and heat loss characteristics. Heat loss was also measured following the completion of the 1999 test series. The results of the heat loss test at the end of the 1999 tests are shown on Figure 2-3.

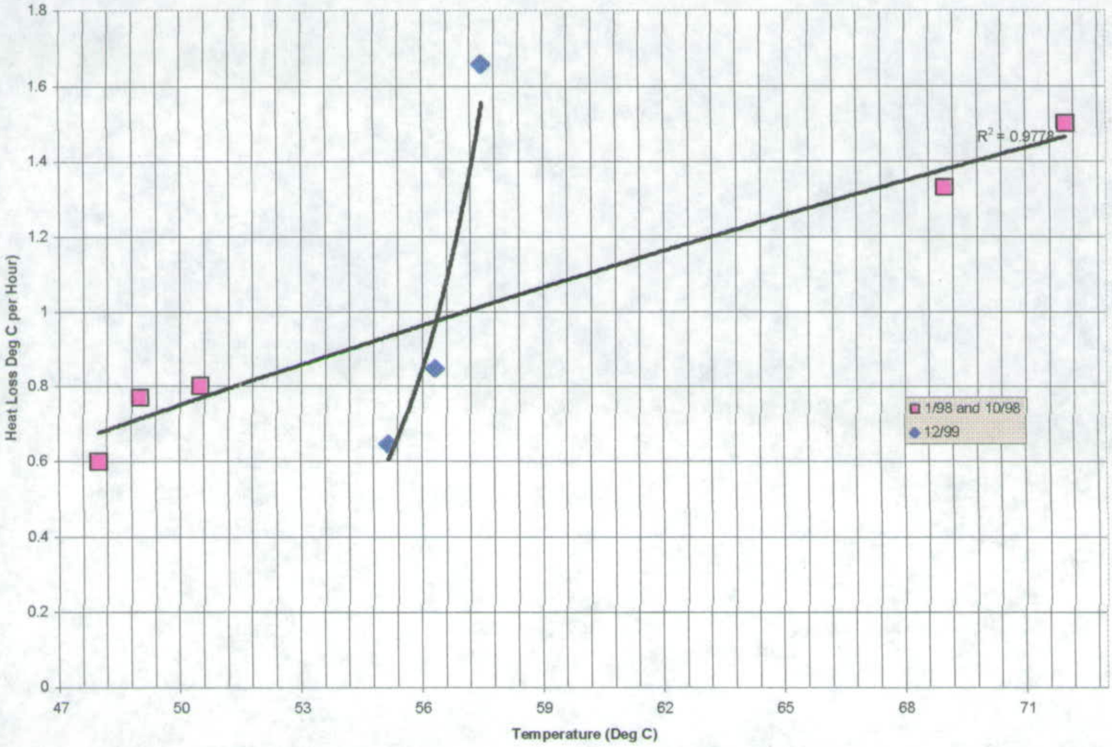
Figure 2-3: Heat Loss vs Temperature after 1999 Tests



The plots indicate a significantly different energy loss from various sectors of the reactor. This is attributed to differing adjacent geology. The upper 160 feet is sand and gravel. Within this zone, flowing water was found between 65 and 115 feet. Bedrock extends from 160 feet to the bottom of the reactor. Within this zone, fractured rock was found at 293 feet and water was found at 315 and 335 feet depth. Heat loss would be expected to vary for these surrounding materials. The heat loss to flowing water would be expected to be significantly higher. Calibration (partially inferred due to physical limitations) indicate that two of the temperature sensors drifted during the test period. Figure 2-3 has been adjusted to reflect corrected temperatures.

Figure 2-4 compares the 1999 weighted average temperature decline for the entire reactor with the heat loss data collected prior to the initiation of testing in early 1998. The comparison indicates a rather dramatic change in reactor heat loss characteristics over the test period, with the effect of increasing temperature on heat loss being much more significant at the end of the test compared to the beginning.

Figure 2-4: Comparison of Heat Loss during 1998 and 1999



The use of this heat loss information is discussed in more detail in the section on energy.

2.5.2 Biological Process Startup and Thermophilic Operation

The initial strategy for biological start-up was to pre-heat the subsurface geology to reduce the rate of heat loss from the reactor. Based on preliminary calculations it was believed that the pre-heating would allow the biological activity to self-heat to desired operating temperatures. A similar approach is commonly used with a variety of composting processes and the ATAD process. The subsurface materials around the reactor were pre-heated by circulating hot water (at 70 to 80°C) from a portable propane fired boiler through the reactor.

Heating of the surrounding geology provided an initial heat source for the biological process and reduced heat loss from the process during biological start-up. However, initial heating was found to only support reactor temperatures of 40 to 43°C when feeding solids. To provide additional heat, hot water generated from a portable diesel boiler was pumped in and out of the extraction line which served as a heat exchanger. Using this method, the process achieved auto-thermal conditions (no supplemental heat input) on January 29, 1998. Temperatures declined after a week of auto-thermal operation, possibly due to additional heat loss resulting from the initiation of dewatering at a neighboring construction site. Based on this situation a steam boiler was added to provide supplemental heat to the process prior to proceeding with the testing program. Following installation of the dedicated propane steam boiler it was found that the process would remain in the thermophilic temperature range. Supplemental heat was required throughout the testing program to maintain thermophilic temperatures. Adding steam to heat the feed resulted in dilution of the THS solids. When the new boiler and heat exchanger were added in 1999, this dilution did not occur. The feed to the reactor was therefore thicker during the initial series of 1999 tests.

During start up, biological activity began generating substantial heat within one to two days. The thermophilic population was found to be quick to respond and recover from feed interruption. This characteristic was consistently observed throughout the testing program.

2.5.3 Operating Conditions

The range of operating conditions of interest for the process are summarized on Table 2-3.

Table 2-3: Desired Range of Operating Conditions

Operating Variable	*Operating Range*
Hydraulic Residence Time (Days)	3 to 6
Temperature (°C)	55 to 70
Aeration (scfm)	20 to 80
Feed Solids Content (%)	5 to 7

The original test plan and schedule called for operating at three detention times (three days, four days, and a third to be determined based on the results of the first two tests). Each of the tests were to include a process stabilization period of at least three detention periods (i.e. nine days for a three day hydraulic retention time (HRT). Solids reduction was used to determine process stability. Once the process was stable, a set of intensive monitoring samples were to be collected to document process performance. This planned sequence was interrupted by the previously discussed mechanical, control, and heat balance problems.

Ten planned performance tests have been completed. The first was initiated on April 6, 1998 and monitored operation with a 3-day HRT. The second was initiated on May 4, 1998 and monitored operation with a 4-day HRT. The third was initiated on December 13, 1998 and monitored operation with a 4-day HRT. In all three tests the process was operated to maximize the operating temperature. During these tests, supplemental heat was provided by injecting steam into the feed tank. The boiler capacity and temperature control in the feed tank limited the process to the 53 to 58°C operating temperature in the initial two tests. During the first two tests, the reactor temperature varied significantly due to interruption of feed and steam supply. Temperatures were more stable in the third test with the daily averages ranging from 55 to 57°C.

Following additional improvements to the demonstration facility an additional series of seven tests were completed between August and December of 1999. The focus of the performance analysis in this report focuses on the third test of the initial series and the entire 1999 test series.

Table 2-4 provides a summary of the operating history of the demonstration facility.

Table 2-4: Summary Operating History

	Operating Conditions		
Dates	*Supplemental Heat*	*HRT (days)*	*Temperature (°C)*
12/18/97 - 1/14/98	Portable boiler	Clean water only	75 to 80
1/15-20/98	None	3	42
1/21-29/98	Steam cleaner	3	54
4/6-8/98	Steam boiler	3	55
5/4-6/98	Steam boiler	4	53
12/13-17/98	Steam boiler	4.2	56
7/29-8/11/99	Hot water boiler	2	67
8/12/-9/30/99	Hot water boiler	4	67
10/7-19/99	Hot water boiler	4	63
10/20-10/27/99	Hot water boiler	4	56
10/28-11/18/99	Hot water boiler	4	56
11/19-12/1/99	Hot water boiler	3.4	56
12/2-21/99	Hot water boiler	5.5	62
12/21/99-2/1/2000	Hot water boiler	Clean water only	60

2.5.4 Performance Data Presentation and Summary

Performance data for the demonstration operations is presented separately for the 1998 and 1999 test series. Major changes in the demonstration facility were completed between these tests. The ability to monitor and record the liquid levels in the process tanks was improved from very limited to continuous between these test periods. The method of providing supplemental heat was also changed from steam heating the feed, which resulted in dilution to hot water heat exchange that did not dilute the feed. These major differences are the basis for presenting the data separately.

2.5.4.1 1998 Test Series

The average solids and organic loading conditions for the initial three 1998 tests periods are provided in Table 2-5.

Table 2-5: Intensive Sampling Periods and Operating Conditions

	Test 1 – 4/6-8/98	*Test 2 – 5/4-6/98*	*Test 3 – 12/13-17/98*
Hydraulic detention time (days)	3*(1)*	4*(1)*	4
Total Solid Loading ppd (kg/d)	1083 (491)	812 (368)	834 (378)
Volatile Solids Load ppd (kg/d)	856 (388)	642 (291)	653 (296)
Ppd/cf	1.2	0.9	0.9
COD Load ppd (kg/d)	ND	ND	930

ND = No Data
(1) = Based on 70 cf batches

Performance data for the 1998 test series are provided on Table 2-6.

Table 2-6: Results of Operation at 3 and 4 Day HRT

	3 Day HRT 4/6-4/8/98	*4 Day HRT 5/4-5/6/98*	*4 Day HRT 12/13-12/17/98*
Total Solids (%)			
THS	6.2%	6.2%	7.4%
Product	4.6%	4.4%	4.7 %
Volatile Solids (%)			
THS	4.9%	4.9%	5.8%
Product	3.4%	3.3%	3.4 %
Removal	32%	33%	41%
Ammonia (mg/L)			
THS (mg/L)	345	480	736
Product (mg/L)	1,520	1,730	2,500
Mg/kg. dry	33,000	39,300	53,200
TKN			
THS (mg/L)	3,521	3,218	5,006
Product (mg/L)	3,287	3,130	4,330
Mg/kg. dry	71,500	71,100	92,100
Organic N (mg/L)			
THS (mg/L)	3,180	2,740	4,320
Product (mg/L)	1,770	1,400	1,830
Mg/kg. dry	38,500	31,800	38,900
Removal	44%	49%	58%
Fats, Oils & Grease (FOG) (mg/kg)			
THS			32,000
Product			4,300
% Removal			92%
Chemical Oxygen Demand (%)			
THS			8.2%
Product			4.2%
% Removal			48%
pH			
THS	5.9	6.2	6.4
Product	7.9	8.0	7.8
Alkalinity			
THS	1,800	4,000	2,410
Product	8,750	6,900	7,340
Temperature °C (average)			
Upper	53-57 (55)	53-54 (53)	55-57 (56)
Lower	54-58 (57)	53-55 (54)	57-59 (58)
Bottom	52-56 (55)	51-54 (53)	55-57 (56)

Mechanical, electrical, and control problems with the demonstration facility and the feed pump delivering thickened solids inhibited stable operation of the facility during these tests. Regardless of these potential upset inducing conditions, the process functioned consistently and was able to provide over 42 percent Volatile solids reduction in the last test. Other basic findings from these tests include:

- Removal (degradation fractions) of COD (chemically oxidized organics), Organic N (a measure of protein) and FOG (fats) were higher than volatile solids reductions.
- The pH increased from about six to about eight through the process.
- Ammonia was released in the process.
- Alkalinity increased through the process.

2.5.4.2 1999 Test Series

Table 2-7 is a summary of the operating periods and conditions for the 1999 tests.

Table 2-7: 1999 Test Period Test Conditions

Condition Start Date	*Condition End Date*	*System Stability (# of days)*	*Length of sampling period (# of days)*	*HRT (Days)*	*Reactor Temp. (°C)*	*Aeration Rate scfm*
12/1/98	12/16/98	16	4	4.7	56	56.2
7/29/99	8/11/99	12	3	2	67	76.5
8/12/99	9/30/99	16	3	4	67	80
10/7/99	10/19/99	12	2	4	63	80
10/20/99	10/27/99	7	3	4	56	60
10/28/99	11/18/99	22	12	4	56	36
11/19/99	12/1/99	12	2	3.38	56	36
12/2/99	12/21/99	19	4	5.45	62	30

A range of detention times and operating temperatures were evaluated. Based on previous test results, the last two tests were conducted with pre-diluted feed to the reactor designed to reduce the viscosity of the reactor liquid with the intent of improving oxygen transfer.

Table 2-8 shows the average concentrations of critical parameters in the thickened solids and product from the VERTAD™ process.

Table 2-8: Average Values for Feed Solids and Product Parameters During 1999 Tests

Thickened Solids (THS)								
Test Reference	*TS Mg/l*	*TVS Mg/l*	*COD Mg/l*	*TKN Mg/l*	*NH3 Mg/l*	*FOG Mg/kg*	*Alkalinity Mg/l*	*pH*
12/98 Trial #3 (4d HRT, 56C)	73638	57596	82171	5006	736	32004	2410	6.4
8/99 Trial (2dHRT, 65C)	65745	52868	77305	3714	573	25700	2760	6.0
9/99 Trial (4d HRT, 65C)	72906	60920	80116	4240	512	25900	2733	6.3
10/99 Trial (4d HRT, 61C)	64131	52583	79377	4102	588	--	3240	6.3
10/99 Trial (4d HRT, 56C)	69183	56842	--	--	--	--	--	--
11/99 Trial (4d HRT, 56C)	65936	53197	88374	3745	513	--	4080	6.3
11/99 Trial (4d HRT, 56C, Dil'd)	72444	56599	81293	3950	547	--	4080	6.3
12/99 Trial (6d HRT, 61C, Dil'd)	69634	55349	77450	3904	456	55400	3240	6.3
VERTAD™ Product								
Test Reference	*TS Mg/l*	*TVS Mg/l*	*COD Mg/l*	*TKN Mg/l*	*NH3 Mg/l*	*FOG Mg/kg*	*Alkalinity Mg/l*	*pH*
12/98 Trial #3 (rd HRT,56C)	46919	33623	42326	4329	2500	4327	7338	8.1
8/99 Trial (2d HRT, 65C)	56815	42850	37883	4101	1544	6000	6420	8.8
9/99 Trial (4d HRT 65C)	52602	40324	58879	4044	1900	5710	6607	8.8
10/99 Trial (4d HRT, 61C)	52289	40163	59144	3914	1794	--	5880	8.0
10/99 Trial (4d HRT, 56C)	52887	40667	--	--	--	--	--	--
11/99 Trial (4d HRT, 56C)	48310	36287	54483	3789	1836	--	6915	8.1
11/99 Trial (4d HRT, 56C, Dil'd)	35183	25393	40238	3113	1633	--	6780	8.46
12/99 Trial (6d HRT, 61C, Dil'd)	31734	22943	33144	2332	1064	11400	4900	8.1

Table 2-9 provides the complete daily feed and product quality information for the August and September 1999 tests.

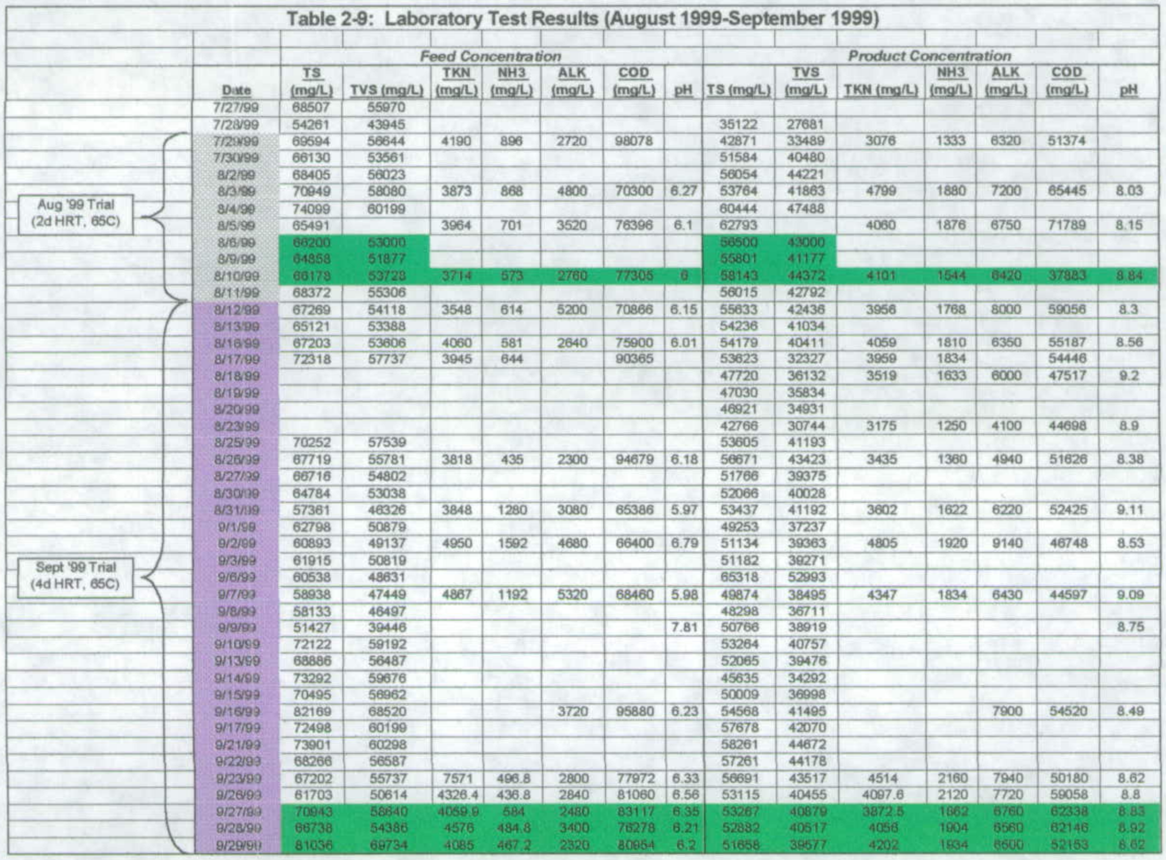


Table 2-9: Laboratory Test Results (August 1999-September 1999)

		Feed Concentration							Product Concentration						
	Date	TS (mg/L)	TVS (mg/L)	TKN (mg/L)	NH3 (mg/L)	ALK (mg/L)	COD (mg/L)	pH	TS (mg/L)	TVS (mg/L)	TKN (mg/L)	NH3 (mg/L)	ALK (mg/L)	COD (mg/L)	pH
	7/27/99	68507	55970												
	7/28/99	54261	43945						35122	27681					
Aug '99 Trial (2d HRT, 65C)	7/29/99	69594	56644	4190	896	2720	98078		42871	33489	3076	1333	6320	51374	
	7/30/99	66130	53561						51584	40480					
	8/2/99	68405	56023						56054	44221					
	8/3/99	70949	58080	3873	868	4800	70300	6.27	53764	41863	4799	1880	7200	65445	8.03
	8/4/99	74099	60199						60444	47488					
	8/5/99	65491		3964	701	3520	76396	6.1	62793		4060	1876	6750	71789	8.15
	8/6/99	66200	53000						56500	43000					
	8/9/99	64858	51877						55801	41177					
	8/10/99	66178	53728	3714	573	2760	77305	6	58143	44372	4101	1544	6420	37883	8.84
	8/11/99	68372	55306						56015	42792					
Sept '99 Trial (4d HRT, 65C)	8/12/99	67269	54118	3548	614	5200	70866	6.15	55633	42436	3956	1768	8000	59056	8.3
	8/13/99	65121	53388						54236	41034					
	8/16/99	67203	53606	4060	581	2640	75900	6.01	54179	40411	4059	1810	6350	55187	8.56
	8/17/99	72318	57737	3945	644		90365		53623	32327	3959	1834		54446	
	8/18/99								47720	36132	3519	1633	6000	47517	9.2
	8/19/99								47030	35834					
	8/20/99								46921	34931					
	8/23/99								42766	30744	3175	1250	4100	44698	8.9
	8/25/99	70252	57539						53605	41193					
	8/26/99	67719	55781	3818	435	2300	94679	6.18	56671	43423	3435	1360	4940	51626	8.38
	8/27/99	66716	54802						51766	39375					
	8/30/99	64784	53038						52066	40028					
	8/31/99	57361	46326	3848	1280	3080	65386	5.97	53437	41192	3602	1622	6220	52425	9.11
	9/1/99	62798	50879						49253	37237					
	9/2/99	60893	49137	4950	1592	4680	66400	6.79	51134	39363	4805	1920	9140	46748	8.53
	9/3/99	61915	50819						51182	39271					
	9/6/99	60538	48631						65318	52993					
	9/7/99	58938	47449	4867	1192	5320	68460	5.98	49874	38495	4347	1834	6430	44597	9.09
	9/8/99	58133	46497						48298	36711					
	9/9/99	51427	39446					7.81	50766	38919					8.75
	9/10/99	72122	59192						53264	40757					
	9/13/99	68886	56487						52065	39476					
	9/14/99	73292	59676						45635	34292					
	9/15/99	70495	56962						50009	36998					
	9/16/99	82169	68520			3720	95880	6.23	54568	41495			7900	54520	8.49
	9/17/99	72498	60199						57678	42070					
	9/21/99	73901	60298						58261	44672					
	9/22/99	68266	56587						57261	44178					
	9/23/99	67202	55737	7571	496.8	2800	77972	6.33	56691	43517	4514	2160	7940	50180	8.62
	9/26/99	61703	50614	4326.4	436.8	2840	81060	6.56	53115	40455	4097.6	2120	7720	59058	8.8
	9/27/99	70943	58640	4059.9	584	2480	83117	6.35	53267	40879	3872.5	1862	6760	62338	8.83
	9/28/99	66738	54386	4576	484.8	3400	76278	6.21	52882	40517	4056	1904	6560	62146	8.92
	9/29/99	81036	69734	4085	467.2	2320	80954	6.2	51658	39577	4202	1934	6600	52153	8.62

Note: Shaded color legend:

Green-performance monitoring period
Gray and purple mark changed operating conditions

Table 2-10 provides the complete daily feed and product quality information for the October through December 1999 tests.

Table 2-10: Laboratory Test Results (October 1999-December 1999)

		Feed Concentration							Product Concentration						
	Date	TS (mg/L)	TVS (mg/L)	TKN (mg/L)	NH3 (mg/L)	ALK (mg/L)	COD (mg/L)	pH	TS (mg/L)	TVS (mg/L)	TKN (mg/L)	NH3 (mg/L)	ALK (mg/L)	COD (mg/L)	pH
Oct '99 Trial (4d HRT, 61C)	10/7/99	62466	51693	4331	536	2960	87059	6.26	57760	44712	3914	1566	5960	69412	7.85
	10/8/99	68861	55282						56512	43591					
	10/11/99	64103	52454						54697	41770					
	10/12/99	65561	52445	4027	718	3200	75097	6.18	54599	41929	3831	1524	7400	58366	8.05
	10/13/99	59874	48874						55107						
	10/14/99	59693	48676	3818	494	2100	67711	6.26	53219	40556	3664	1622	5780	58708	7.91
	10/15/99	62552	50866						53689	41050					
	10/18/99	62573	51261						52512	40397					
	10/19/99	65688	53904	4102	588	3240	79377	6.32	52066	39928	3914	1794	5880	59144	8
Oct '99 Trial (4d HRT, 56C)	10/20/99	66413	54172						52041	39386					
	10/21/99	67394	55155	4414	840	2880	98824	6.34	54202	42256	4310	1692	5810	58824	7.92
	10/22/99	72181	58410						54086	41731					
	10/25/99	67012	55241						53848	41380					
	10/26/99								51978	39857					
	10/27/99	71354	58442						52836	40663					
Nov '99 Trial (4d HRT, 56C)	10/28/99	73231	59816			3300		6.48	44514	34083			6020		8.64
	10/29/99	64644	53192						42164	32546					
	11/1/99	64423	53200						46261	35325					
	11/2/99	63205	51651	4089	536	4480	81792	6.25	46974	35822	3789	1768	7510	52224	8.18
	11/3/99	64814	53348						47687	36334					
	11/4/99	66633	54450						47363	36581					
	11/5/99	62849	51125						47825	36294					
	11/8/99	64513	53023						47846	36281					
	11/9/99	67968	55638	3400	490	3680	94956	6.25	47600	36201	3789	1904	6320	56742	8.07
	11/10/99	68633	55867						48068	38346					
	11/12/99	68405	53241						51100	38626					
	11/15/99	65032	51785						48589	35055					
	11/16/99	64150	51098	3448	440	2320	73958	6.16	50444	36292	3664	1934	8840	52882	7.99
	11/17/99	65942	52156						47959	34681					
	11/18/99	69084	54976						48264	34929					
Nov '99 Trial (4d HRT, 56C, Dil'd)	11/19/99	68071	54149						49666	36491					
	11/22/99	72020	56773						43642	30388					
	11/23/99								41531	29051	3477	1862	5640	43112	8.62
	11/24/99	73088	58295						39200	27403					
	11/26/99	71456							44701	30983					
	11/29/99								40571	29501					
	11/30/99	76682							38977	28439	3113	1633	6780	40238	8.46
	12/1/99								36017	25834					
	12/2/99	68994	54728						34348	24952					
Dec '99 Trial (6d HRT, 61C, Dil'd)	12/3/99	66997	53780						32515	23290					
	12/6/99	67897	54210						26481	19102					
	12/7/99	72461	56723						29279	21332			6560	28789	8.57
	12/8/99	71158	57549						28520	20723					
	12/9/99	70589	56509						29841	21675					
	12/10/99	72026	57705						33603	24380					
	12/13/99	71577	56845						33816	24682					
	12/14/99	68193	54843						40217	29116					
	12/15/99								39270	27845					
	12/16/99	75340	54350	3876.7	429	4000	80645	6.05	42388	31051	2789.9	1160	5000	37279	7.71
	12/17/99	66560	50691	3502.3	458	2600	80620	6.23	33247	24139	1435.2	1040	4600	32558	8.14
	12/19/99	67456	53774	4120.3	481.6	3920	75680	6.03	35727	25281	2707	1072	5100	34813	7.87
	12/20/99	68420	54640	4089	428.8	3200	76051	6.6	35092	25174	2852.3	1079	5000	32061	8.23

Note: After 11/19/99, dilution water was added to the feed and accounted for in the mass balance.

Table 2-11 provides a statistical summary of the quality data throughout the 1999 test period. The THS and product information summarizes data for the entire period. The diluted feed data is included on the last two tests in November and December.

Table 2-11: Statistical Summary of Feed and Product Quality

	TS *Mg/l*	*TVS* *Mg/l*	*TKN* *Mg/l*	*NH_3* *Mg/l*	*ALK* *Mg/l*	*COD* *Mg/l*	*pH*
Thickened Solids							
Avg.	67,317	54,456	4,162	649	3,325	80,421	6.3
Median	67,394	54,418	4,060	537	3,200	78,675	6.2
SD	4,996	4,292	751	282	869	9,617	0.3
SD/Avg. %	7.4	7.9	17.0	43.4	26.1	12.0	5.4
Max	82,169	69,734	7,571	1,592	5,320	98,824	7.81
Min	51,427	39,446	3,400	429	2,100	65.386	5.97
No	91	88	29	29	30	30	30
Diluted Feed							
Avg.	42,916	33,111	2,687	946	3,869	50,933	6.4
Median	42,417	31,653	2,904	968	3,800	51,938	6.5
SD	6,869	5,750	620	100	716	3,049	0.3
SD/Avg. %	16.0	17.4	23.1	10.5	18.5	6.0	4.2
Max	62,183	49,645	3,106	1,040	5,000	54,547	6.8
Min	31,715	24,370	1,477	772	2,760	46,409	6.1
No	21	21	6	6	7	7	7
Product							
Avg.	48,723	36,580	3,724	1,660	6,459	51,372	8.4
Median	51,134	38,626	3,873	1,768	6,420	52,425	8.5
SD	8,192	6,871	662	299	1,138	10,658	0.4
SD/Avg. %	16.8	18.8	17.8	18.0	17.6	20.7	4.8
Max	65,318	52,993	4,805	2,160	9,140	71,789	9.2
Min	26,481	19,102	1,435.2	1,040	4,100	28,789	7.7
No	99	97	33	33	35	35	35

Considering the ratio of standard deviation to average, the data indicates that the THS quality was relatively consistent for TS, TVS, COD and pH. TKN and alkalinity were more variable. Ammonia in the THS was more variable, although not as significant a factor compared to the TKN values. The product was less consistent than the THS for TS, TVS and COD. It was more consistent in ammonia content.

A critical analysis procedure was to prepare a mass balance for each of the test periods. An example of this analysis is provided in Table 2-12. The analysis shown is for the final test during which the process was operated with a six-day HRT and a set point temperature of 61°C.

Table 2-12.: Example Mass Balance Calculation for Dec. 1999 6 Day HRT Test

Raw Sludge Input		
THS input	172	in
	286.67	ft3
	2144.41	gal
Cycle Time	70	hr
	4200	min
THS Feedrate	0.511	gpm

Dilution Water Input		
Water input	66.5	in
	110.83	ft3
	829.09	gal
Cycle Time	70	hr
	4200	min
Water Feedrate	0.197	gpm

Evaporative Losses			
		Constants	
	A	B	C
	18.3036	3816.44	-46.13
Feed Tank Temperature	40	C	
Vapour Pressure H2O	54.754213	mmHg	
Compressor Load	32	%	
Aeration Rate	27.84	scfm @68 F	
lb moles air	0.0722291	lbmoles/min	
Total P	760	mm Hg	
P air	705.24579		
lb moles water	0.0056078	lbmoles/min	
	0.1009397	lb/min	
	0.0016176	ft3/min	
	2.3293776	ft3/day	
	0.012	gpm	

Reactor Temperature	61	C
Actual HRT	5.45	days
Average Aeration	22.00	scfm
Average Off-gas	10.50	%
Average OTE	50.0%	

Period for this Mass Balance: Trial Average of stable condition periods
(Fat addition period, and 1 odd grab sample have been omitted.)

Mass Balance		
THS VS	55349	mg/l
Product VS	22943	mg/l
FLOW		
THS+dilution	0.708	gpm
Water (Evap)	0.012	gpm
Product	0.696	gpm
VOLATILE SOLIDS		
VS in	14.14	lb/hr
VS out	7.99	lb/hr
VS Destruction	43.5%	
Actual HRT	5.45	days
THS COD	77450	mg/l
Product COD	33144	mg/l
FLOW		
THS+dilution	0.708	gpm
Water (Evap)	0.012	gpm
Product	0.696	gpm
VOLATILE SOLIDS		
COD in	19.78	lb/hr
COD out	11.54	lb/hr
COD Destruction	41.7%	
Actual HRT	5.45	days
THS Org-N	3448	mg/l
Product Org-N	1268	mg/l
FLOW		
THS+dilution	0.708	gpm
Water (Evap)	0.012	gpm
Product	0.696	gpm
VOLATILE SOLIDS		
Org-N in	0.88	lb/hr
Org-N out	0.44	lb/hr
Org-N Destruction	49.9%	
Actual HRT	5.45	days

Using the mass balance procedure, the performance of the process was determined for each test period. The results are summarized on Table 2-13. For comparison between the 1998 and 1999 test periods, the last test in the 1998 series is included in this data presentation.

Table 2-13: VERTAD™ Digestion Performance Summary

Test Reference	*Net Water Loss (Lb H2O per Lb Sludge Fed)*	*VS Destruction (% Removal)*	*COD Destruction (% Removal)*	*Org-N Destruction (% Removal)*	*FOG Destruction (% Removal)*
12/98 Trial #3 (4d HRT, 56C)	-0.107	40.9	48.0	57.9	91.7
8/99 Trial (2d HRT, 65C)	0.638	20.7	52.0	20.3	–
9/99 Trial (4d HRT, 65C)	2.072	42.2	35.8	49.8	80.8
10/99 Trial (4d HRT, 61C)	2.217	33.9	35.5	47.8	–
10/99 Trial (4d HRT, 56C)	0.331	30.1	–	–	–
11/99 Trial (4d HRT, 56C)	0.197	32.7	39.2	40.4	–
11/99 Trial (4d HRT, 56C, Dil'd)	N/A	42.3	36.3	44.1	–
12/99 Trial (6d HRT, 61C, Dil'd)	N/A	43.5	41.7	49.9	–

The data shows the process providing volatile solids reductions ranging from 21 to 43 percent for various operating conditions. As with the first test series the organic nitrogen and fat degradation rates were consistently higher than the volatile solids reduction rate. However, in this test series, the COD reduction was equally greater and less than the volatile solids reduction.

2.5.5 Process Simplicity and Stability

The biological process was found throughout the testing program to be relatively simple to operate, resistant to upset, and to rapidly recover from disruptions caused by electrical and mechanical system failures. The straightforward process controls consist of providing a supply of food on a relatively *uniform basis and providing air. In a full-scale system, with continuous feed and batch discharge and* manually adjusted aeration to match growth induced loading increases and seasonal fluctuations, the operational controls would be expected to require less operator attention than an anaerobic digestion process. The VERTAD™ process operates well over a range of pH conditions and temperatures. Although the process does not generate gas, it does produce hot water and does not require the extensive gas handling, cleaning, and safety equipment.

The ability of the process to recover quickly from upset conditions was demonstrated on numerous occasions as the result of power outages and failure at the feed system, boiler, or control system. During these occasions, the process was stressed by lack of food, cooling, and aeration. In all situations the process recovered rapidly.

An example is a period between September 3 through 14, 1999. During this 12-day period the system for discharge of product was not functioning properly, which restricts the ability to feed the reactor. The boiler was able to maintain the operating temperature in the mixed zones of the reactor but not in the soak zone which cooled between discharge events. Table 2-14 presents the total solids, volatile solids and chemical oxygen demand for the upset and stabilization periods. The table also shows the same data for the test period sampling that immediately followed the stabilization period. The mass balance for the test period found a VSR for this period of 42 percent (among the best volatile solids reduction performance observed during testing).

Table 2-14: Average THS and Product Quality During and After Upset

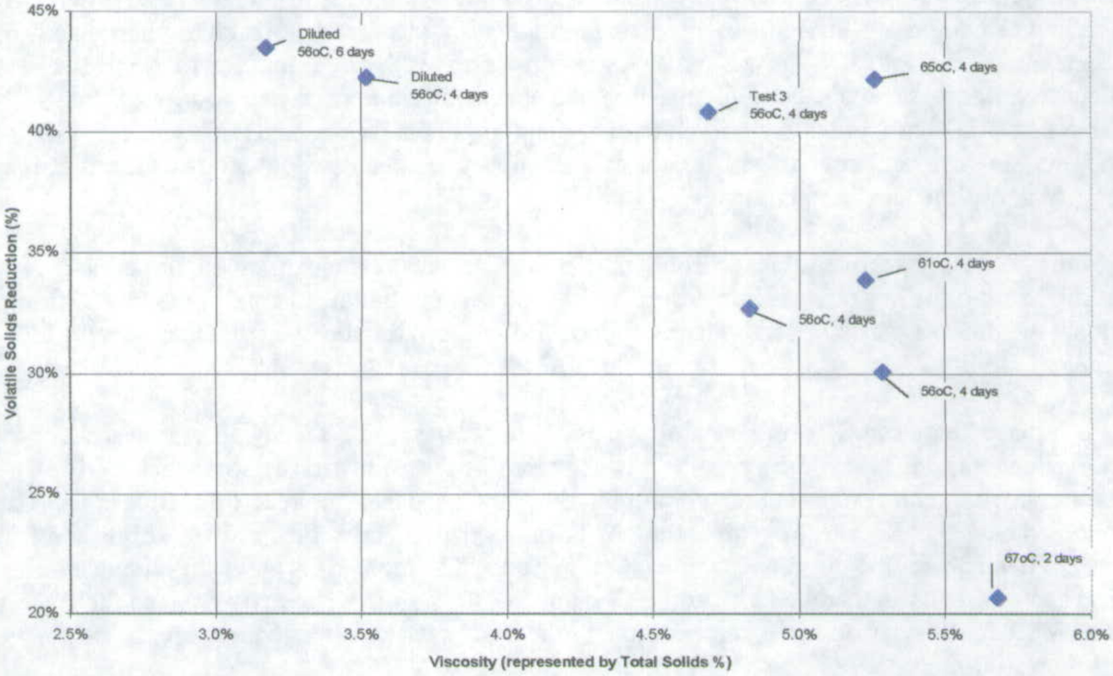
	THS (mg/L)			*Product (mg/L)*		
	TS	*VS*	*COD*	*TS*	*VS*	*COD*
Upset Period (7 days)	63,500	51,100	68,500	52,200	40,200	44,600
Stabilization Period (7 days)	70,900	58,400	85,000	55,400	41,900	54,600
Test period (3 days)	72,900	60,920	80,100	52,600	40,300	62,200

The product data for total and volatile solids are very stable considering the extent of the upset and extreme variation in feed over the three periods. The product COD data appears to indicate a change in performance over the period, which is reflected by the very low 36 percent COD removal reported for the test period.

2.5.6 Residence Time and Feed / Reactor Content

Figure 2-5 summarizes the performance of the demonstration facility for the eight identified test periods using volatile solids reduction and reactor total solids content as the comparative parameters. The figure also shows the primary test variables of detention time and temperature. Examination of the data indicates the following general trends:

Figure 2-5: VS Destruction vs Reactor Viscosity



- An apparent maximum VSR of 44 percent over the operating range of two to six-day detention time.
- A trend toward reduced VSR as the solids content (viscosity) of the reactor liquor increases.
- Considerable performance variability at a four day HRT.
- Possibly improved performance at higher temperatures when the reactor solids are high.

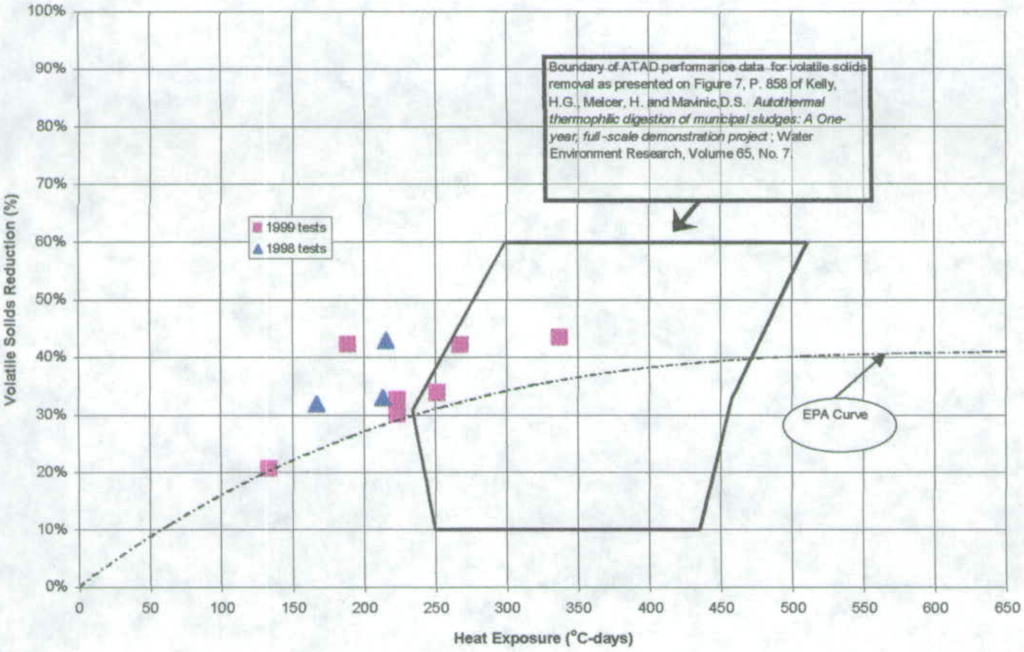
Based on this data, an interim conclusion is that a VERTAD™ reactor operating with a four day HRT, at a temperature of 56°C or greater, and less than 4.5 percent reactor TS, will provide the maximum VSR demonstrated to date.

2.5.7 Vector Attraction Reduction

The selected method of vector attraction reduction (VAR) for the VERTAD™ process is to attain 38 percent reduction in volatile solids. The performance of the VERTAD™ process in comparison to previously published curves for the Autothermal Thermophilic Aerobic Digestion (ATAD) process is provided on Figure 2-6. The comparative parameter in this case is the product of time and temperature. This comparison indicates that the VERTAD™ process typically operates and provides VAR with fewer degree days, but so far has not produced the upper range of VSR measured at ATAD facilities. VERTAD™ has consistently operated at or above the EPA performance curve. The currently used mesophilic anaerobic digestion process would be off of this figure with a VSR of 60 to 70 percent and heat exposure of 800 to 1200 °C days.

Additional discussion of VSR and organic degradation is provided in Section 2.5.12.

Figure 2-6: Volatile Solids Reduction - Comparison of EPA design curve, ATAD demonstration results and VERTAD test results



Boundary of ATAD performance data for volatile solids removal as presented on Figure 7, P. 858 of Kelly, H.G., Melcer, H. and Mavinic,D.S. *Autothermal thermophilic digestion of municipal sludges: A One-year, full-scale demonstration project*; Water Environment Research, Volume 65, No. 7.

2.5.8 Class A Pathogen Control

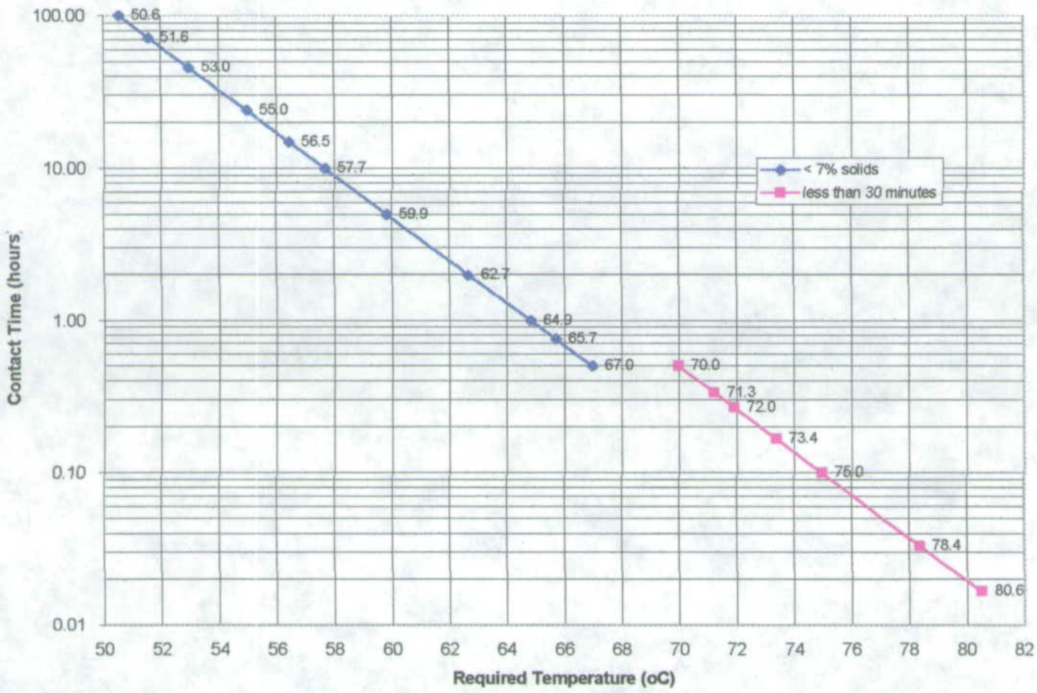
Class A pathogen control requires compliance with three criteria:

1. Fecal coliform and salmonella
2. Vector Attraction Reduction process
3. Pathogen control process

Compliance with each of these Class A and Vector Attraction Reduction requirements is achievable with the VERTAD™ process. Microbial analysis has consistently found that fecal coliform and salmonella levels in VERTAD™ product are below detection limits (0.03 MPN/gm dry wt. and 0.2 MPN/gm dry wt. respectively). The THS processed through the system contained from $4x10^7$ to $3x10^{10}$ feed coliform, and from non-detected to 540 salmonella organisms. The process has provided greater than 40 percent volatile solids reduction in four days detention time at a 56°C operating temperature, provided that the total solids in the reactor is maintained below 4.5 percent. Higher removal rates for organic nitrogen (>45 percent) and fats, oils and grease (>80 percent) indicate rapid degradation of these fractions which primarily produce the objectionable character of un-stabilized wastewater solids.

The selected alternative for attaining Class A pathogen control in the VERTAD™ process is by maintaining temperatures for the required contact time. Time and temperature requirements from the biosolids regulations (40CFR503 and WAC 173-308) are shown on Figure 2-7.

Figure 2-7:
40 CFR 503 Class A Time and Temperature Requirements
for Solids Less Than 7%



A conservative approach to compliance would be to control the batching cycle and temperature such that all fed solids would be exposed to the required temperature between the end of feeding and the beginning of the discharge cycle. In this way the entire reactor contents would be in compliance at the end of each batch. Assuming a four-hour batching cycle and one hour from beginning of discharge to end of feeding, the reactor would need to be maintained at a temperature of 62°C. The soak zone can also be used to provide the required time and temperature. Typically the temperature in the soak zone cools to 0.9 to 1.5°C below the upper zones at the end of the cycle. To operate such that the pathogen requirement is met in the soak zone, the reactor would need to be operated at 64°C in order to assure adequate soak zone temperatures throughout the batch cycle.

While it is believed that the demonstration facility's vertically stacked zone configuration complies with the time and temperature requirement, two variations are available to further assure compliance:

- Installation of a flow restricting physical barrier between the slowly mixed and soak zones.
- Maintain a surface batch contact tank in which the VERTAD™ product is held for the required time at the appropriate temperature.

NORAM is currently working with the EPA to determine the appropriate method of documenting Class A compliance in the VERTAD™ reactor. Section 2.5.15 includes a discussion of reactor mixing and zone separation that are important factors for Class A time and temperature compliance.

2.5.9 Product Suitability for Beneficial Use

VERTAD™ production of a Class A product expands the potential markets for King County Biosolids as a commercial product. Site permitting would no longer be required, so the material could potentially be used without further processing by nurseries, topsoil manufacturers and landscapers. These markets, however, have fairly restrictive quality requirements. Portions of this market are currently accepting the City of Tacoma's Tagro product, which is a mix of Class A anaerobically digested biosolids mixed with sawdust and sand.

The visual and olfactory character of the solids are very important for these markets. Dewatered VERTAD™ product has been found to have a different appearance than the current anaerobic product. The VERTAD™ product has more of a gray-brown color and a fibrous texture, like wet paper. This product may be useable for making a product similar to Tagro or for use in topsoil manufacturing. Minimal additional processing by composting or drying could produce a high value organic product that could be sold locally. Local sales of these types of products would dramatically reduce the cost of biosolids management for the County by decreasing transportation costs.

Nitrogen content is a factor in determining the value of the product. The performance data indicates that the product has a lower organic nitrogen content (3.2 to 3.9 percent as N) compared to typical South Plant anaerobically digested solids (4.1 to 7.6 percent as N). No data is available for nitrogen content of a dewatered VERTAD™ product.

2.5.10 Anaerobic Linking

Thermophilic aerobic digestion has been used in a dual digestion process where pure oxygen activated sludge systems are in use. The pure oxygen stimulates the development of thermophilic temperatures in very short duration reactors (about 24 hours). These reactors were normally followed by mesophilic anaerobic digestion. The City of Tacoma operates this process. The potential for a similar linked aerobic thermophilic and mesophilic anaerobic digestion system was evaluated by the University of Washington (Ref. 15). The product from the VERTAD™ reactor operating at a four day HRT was used as feed for bench scale anaerobic digestion tests. The results of these tests are provided on Table 2-15.

Table 2-15: Comparison of Anaerobic Control with Combined System Performance

	11 day SRT Anaerobic Control	*15 day Anaerobic with VERTAD™*	*11 day Anaerobic with VERTAD™*
Solids Retention Time (days)			
VERTAD™	0	4	4
Anaerobic	11	15	11
Total	11	19	15
Volatile Solids Reduction (%)			
VERTAD™	0	33	33
Anaerobic	52	49	45
Total	52	66	63
Anaerobic Gas Production			
Liters Methane / day	2.8	2.0	2.5
Liters Methane / gram COD removed	0.51	0.39	0.36

The data indicates that following VERTAD™ with anaerobic digestion provides significant additional reduction in volatile solids with the production of significant gas. The additional four days of anaerobic digestion provided by the longest SRT test did not produce greatly improved VSR. The higher methane production per weight of COD removed for the anaerobic control seems to indicate that the VERTAD™ process has selectively removed fats from the feed, or the anaerobic control feed contained a higher fat fraction than the reactor fed VERTAD™ product. The complete report for this evaluation is included as Appendix A.

The reverse configuration of anaerobic digestion preceding VERTAD™ was not tested because of physical restriction of available test equipment. The potential differences between these two approaches are discussed in Section 3.

2.5.11 Thickening/Dewatering/Centridry

The ability to thicken, dewater and dry the VERTAD™ product is a critical factor in determining the cost effectiveness of the process. Thickening is of value for reducing the volume of material that is passed on to a linked anaerobic digestion process. Thickening may also remove dissolved compounds which increase the polymer required for dewatering (Ref. 7). The effectiveness of dewatering and drying determines the cost of transporting the product to biosolids use sites.

The VERTAD™ process provides the unique opportunity to float thicken product by adding chemicals to release soluble carbon dioxide as fine gas bubbles. This option is available only when drawing the product directly from the high-pressure lower zone of the reactor.

Flotation thickening tests were performed on VERTAD™ samples by adding 93 percent sulfuric acid or alum to a freshly withdrawn sample, and allowing it to float thicken over a period of 1-2 hours. Acidified and alum conditioned samples float thickened to approximately the same concentration. Sulfuric acid would be the preferred chemical because it costs a great deal less than alum. The mechanism for flotation is:

- Reactor contents are under pressure, and at 60°C or below, ammonia is present in a stable form (as ammonium bicarbonate)
- When a sample is brought up from the depths of the reactor to the surface, the sample depressurizes, and some CO2 is released (most of the CO2 is still dissolved though, and the ammonium bicarb is still stable, the sample is supersaturated)
- Acidifing the sample (with sulfuric acid, or alum) to approximately pH 5, releases the CO2 as small bubbles which attach to biosolids particles, floating the biosolids into a compact blanket. (Ammonium sulfate is formed in this substitution reaction).
- Float thickened solids were between 8-12 percent total solids in the tests (for alum and sulfuric acid)

In alternatives involving dual digestion, the limit in the design is the internal solids concentration in the anaerobic digester (kept at a max of 4.5 percent). So despite the fact that the concentration of the float thickened solids can be up to 12 percent, significantly reducing downstream tankage and dewatering equipment, approximately 7 percent is the maximum allowable float concentration to stay within the digester constraints. In straight VERTAD™ options, a 10 percent float thickness can be used (this is assumed to be quite conservative). A 95 percent capture efficiency is used for the flotation cell (this approximates what was observed in the subnatant).

The operating temperature of the reactor influences the performance of flotation thickening. Somewhere above 60°C, ammonium bicarbonate is unstable and dissociates. This results in more ammonia in the off-gas (which has to be recaptured in a biofilter) and less ammonium bicarbonate in solution for the acidification/float separation stage that follows the VERTAD™. Of the ammonia and CO_2 liberated from this dissociation, the ammonia re-adsorbs into solution much more readily, raising the pH from approximately 8 to 8.8. In terms of design, this means an increased sulfuric acid requirement in the flotation stage that follows to depress the pH to the point where CO_2 is released (approximately pH 5).

Laboratory dewatering tests were completed for VERTAD™ and anaerobically digested product by Andritz. Table 2-16 presents the results of the tests.

Table 2-16: Dewatering of VERTAD™ and Anaerobic Digestion Products

	Anaerobic	*VERTAD™*		
	Digested	*Direct*	*Acid Floated**	*Alum Floated*
Initial Total Solids (%)	3.50	3.64	8.77	6.82
Initial Suspended Solids (%)	3.43	3.45	8.68	6.72
Initial Ash Content (%)	37	27	20	30
Capillary Suction Time (sec)	237	561	318	210
Screen +100 mesh (% of SS)	8	25	31	25
100x230 fraction (% of SS)	9	5	3	4
Solids Capture (% of SS)	95	96	99.5	98
Cake Solids (%TS)	12-14	31-34	31-34	27-30
Neat Polymer Dosage (lb/ton SS)	20	38	14	14
			**pH adjusted*	

Based on tests of cooled products by Andritz Ruthner, Inc. using Ciba 757 polymer (Ref. 9)

The differing character of the anaerobic and VERTAD™ products are apparent from the ash content and screened solids fractions. The VERTAD™ product contains less ash and larger fraction of small particles relative to the anaerobic product. The VERTAD™ product has a higher capillary suction time, but it is reduced to that of the anaerobic product in the thickening process. As with previous tests, the VERTAD™ product dewatered to a very high cake solids compared to anaerobic product, but the polymer demand was almost twice as great. The impact of flotation thickening on dewatering performance was quite impressive. The solids capture increased and the polymer demand was less than for anaerobic product while maintaining the high solids content in the cake. There is no definitive data on the total solids capture through both the flotation and dewatering processes.

2.5.12 Microbial Degradation of Organics

2.5.12.1 Differential Degradation of Fats

Constituent (fats, carbohydrates, and proteins) degradation is an important design factor. As we all know from food product labeling, eating fat provides more calories per gram consumed than eating carbohydrates or protein. This fact is due to the chemical structure of these biochemically active compounds. Fats are basically long chain hydrocarbons with a fatty acid molecule attached at one end. As such, the potential for oxidation is maximized. Available energy per unit mass is similar to fuel oil and gasoline. Degradation of fat requires larger quantities of oxygen for oxidation and produces large quantities of methane during anaerobic reduction compared to protein and carbohydrates. Proteins and carbohydrates are already partially oxidized which reduces the potential energy (as heat or methane) available from degradation. Conversely, in aerobic systems this partial oxidation reduces the oxygen needed for degradation.

Figure 2-8 compares the oxygen required for aerobic degradation for typical fat, carbohydrate and protein compounds. The oxygen requirement for fat is about 2.5 times that of carbohydrate and protein. This figure also compares the degradation products from each. Only protein releases significant ammonia. Fats release almost twice the carbon dioxide and energy. Proteins release the least water while fats release the most.

Figure 2-8: Aerobic Process
Oxygen Consumption and Degradation Products
for Common Organics

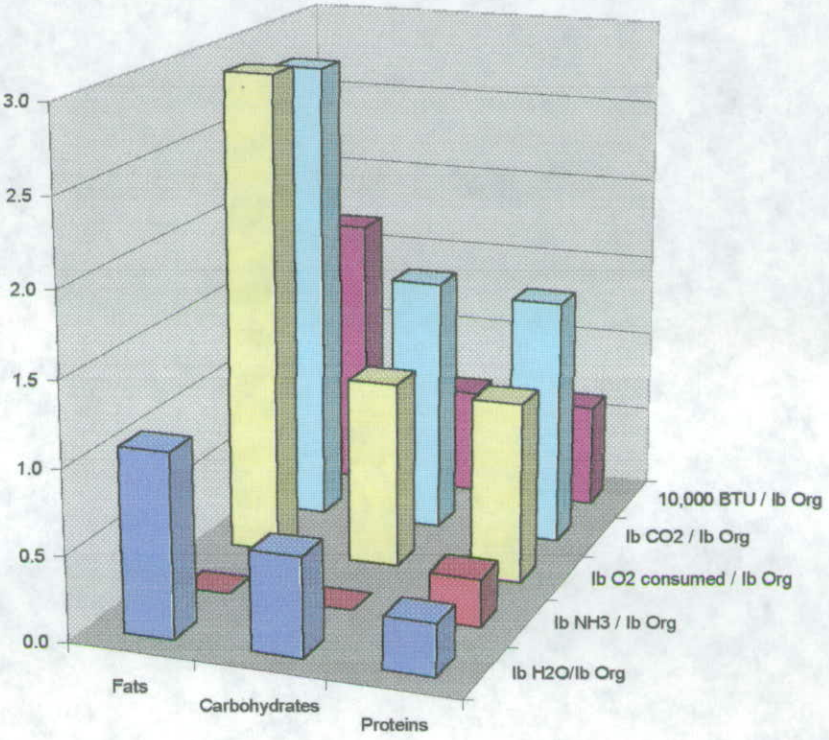
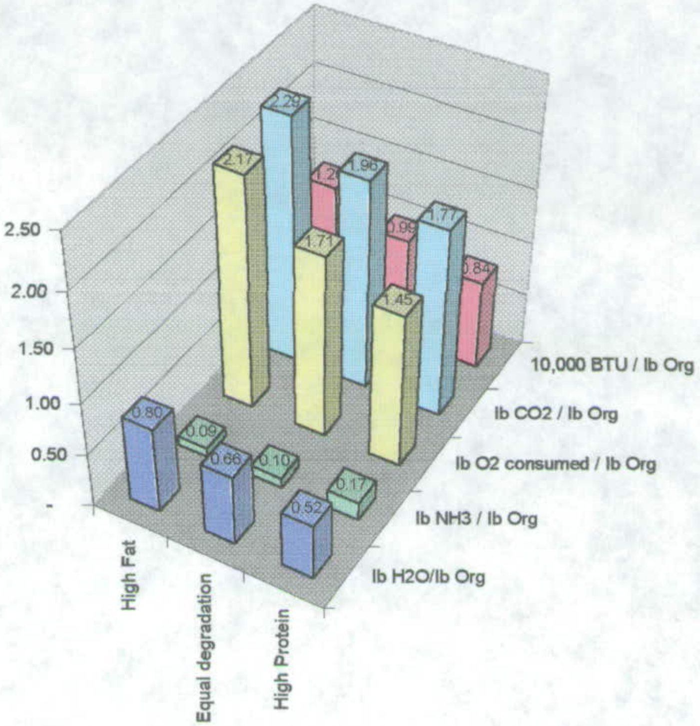


Figure 2-9 shows the effect on these same products and oxygen demand for three constituent mixes. The high fat mix assumes that the degraded organics are 60 percent fat, 10 percent carbohydrate and 30 percent protein by weight. The equal degradation assumes an equal weight of each constituent is oxidized. The high protein mix assumes that 60 percent of the degraded organics is protein together with 20 percent each of fat and carbohydrate being degraded. The figure shows the values for each parameter for each mix. Based on this comparison, it is obvious that differences in feed character or specific process preference for one type of constituent could significantly affect the oxygen demand, energy release and balance between carbon dioxide and ammonia release.

Figure 2-9: Aerobic Degradation Consumption and Products
Impact of High Percentage Fat and Protein Degradation



The most comprehensive evaluation of fats, carbohydrates and protein was completed during the final 1998 test normally referred to as Test 3. Table 2-17 compares the chemical characteristics of typical organic compounds. The table also shows the breakdown characteristics for Test 3 based on measured total, protein, and fat content. Carbohydrates are calculated by difference. The feed to the reactor had a very low fat content of 4 percent compared to a more typical sludge characterization with 10 to 40 percent fat (Ref. 11). Replacement of fat in a typical wastewater solid with carbohydrate or protein represents a reduction in potential energy released per pound of volatile solids degraded from about 12,000 to about 7,300 or a 40 percent decrease.

Table 2-17: Chemical Composition of Primary Organic Compounds

	Chemical Formulas				*By Product Ratios*			*Energy Potential*[1]
	C	**H**	**O**	**N**	**H_2O/Org**	**NH_3/Org**	**Co_2/Org**	**BTU/lb.**
				Typical Compounds				
Lysine (cadaverine)	5	12	2	2	0.41	0.26	1.7	7,700
Sucrose	12	22	11	0	0.58	0	1.5	6,500
Oils	50	90	6	0	1.03	0	2.8	16,400
				Test 3 Breakdown				
Proteins (64%)	33.3	83.1	16.6	16.6				
Carbohydrates (27%)	13	23.8	11.9	0				
Fats (9%)	3.75	6.75	0.45	0				
Test 3 Composite	*50*	*114*	*29*	*17*	*0.51*	*0.20*	*1.6*	*7,300*

[1] Based on 5,780 BTU/lb. O_2 used for oxidation

The VERTAD™ test reactor as configured at the beginning of Test 3 could be treated as a large respirometer. All feed and product streams could be monitored to provide mass balances of the critical components of the biological process (carbon compounds, oxygen, solids, and water). In addition, characterization of heat loss from the reactor allowed estimates of the heat balance for the exothermic degradation of organic compounds to be made.

A primary objective of the VERTAD™ process is to stabilize and reduce the quantity of solids to be dewatered, hauled, and utilized. The standard measure of stabilization is volatile solids reduction. The state and national biosolids regulations require 38 percent volatile solids reduction for typical digestion processes. To determine the volatile solids reduction a mass balance around the entire process was completed using measured densities and levels in the feed and head tanks through batch feeding events. Similar calculation procedures were used to develop mass balance information for chemical oxygen demand (COD), organic nitrogen and fats, oils, and grease. Samples were collected every six hours and composited to make a daily sample for analysis. The results of the mass balance analyses are provided on Table 2-18.

Table 2-18: Organic Matter Degradation (as % removal)

Date	*VS Destruction*	*COD Destruction*	*Org-N Destruction* (1)	*Fats, Oils, and Grease*
12/13/98	41.7%	50.5%	60.7%	94.8%
12/14/98	43.9%	51.4%	50.9%	ND
12/15/98	42.5%	46.9%	60.1%	90.3%
12/16/98	43.7%	49.5%	59.2%	88.4%

(1) Mass Balance using measured densities.

The measured quantities of the three residual organic compounds, fats and oils, carbohydrates, and protein can be estimated using the results of the volatile solids, fats, oils, and grease (FOG) and organic nitrogen analyses. Table 2-19 provides the average pounds of each compound type fed to the reactor daily during the third test. The percent distribution indicates relatively small amounts of fats and oils and equal quantities of protein and carbohydrates. This fat content is unusually low for raw wastewater solids and should be verified in future tests.

Table 2-19: Feed Volatile Solids Distribution for Test 3

	Pounds per day	*% of Total*
Total Volatile Solids	638	
Fats, Oils, and Grease	26	4%
Protein[1]	295	46%
Carbohydrates[2]	316	50%

[1] Assumes 6.25 lb. Protein/lb. Org N per 21CFR 101.9
[2] By difference

Table 2-20 shows the measured degradation of these compound groups in the VERTAD™ reactor. The percent of total indicates that the proteins were the primary compound being degraded during Test 3. Carbohydrates were not degraded as effectively as protein. Although almost all of the fats and oils were degraded, they comprise a small fraction of the organics and therefore represent a small fraction of the amount of organics degraded. This is a significant finding relative to energy release because the fats and oils release about 2.5 times as much energy per pound of material degraded compared to proteins and carbohydrates. The significant residual carbohydrates may be related to an observed papery texture of the dewatered product. This may reflect limited degradation of cellulose and lignin, which would be included as carbohydrates in this characterization.

Table 2-20: Volatile Solids Degradation During Test 3

	Pounds Per Day Degraded	*% of VS Reduced*	*% Energy Release and O_2 Demand*
Total Volatile Solids	265		
Fats, Oil, and Grease	24	9	20
Protein[1]	169	64	56
Carbohydrates[2]	73	27	24

[1] Assumes 6.25 lb. Protein/lb. Org N per 21CFR 101.9
[2] By difference

The COD reduction attained by the VERTAD™ process is higher than the VS reduction. This is consistent with results for other ATAD processes (Ref. 5). A possible explanation is that ATAD™'s and VERTAD™ preferentially degrade fats which increases the oxygen demand of the degraded fraction per mass degraded. This would produce a higher removal of COD compared to volatile solids.

Although no systematic comparative data was collected during the 1999 tests, an examination of the data provides some useful insight. Table 2-21shows statistics for the ratios of COD/VS and Organic nitrogen/VS, and FOG/VS for the thickened feed solids and the VERTAD™ product.

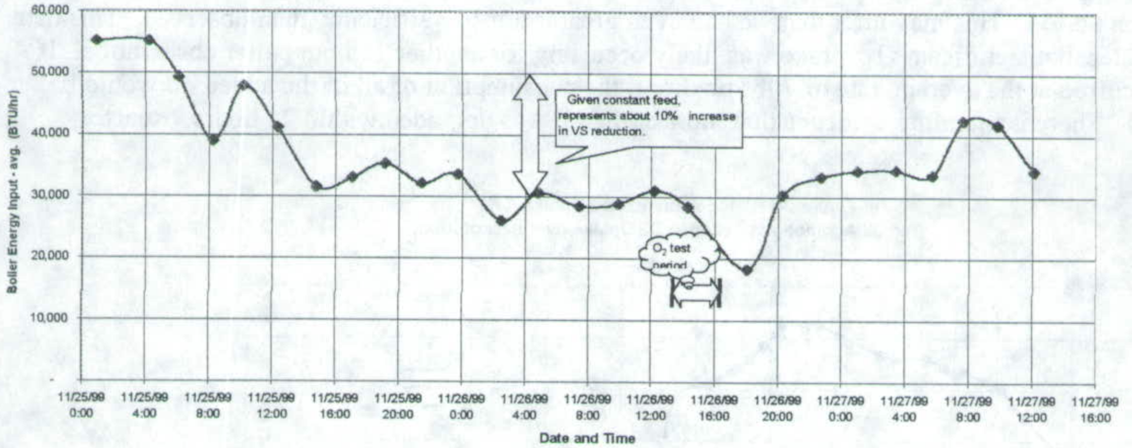
Table 2-21: Volatile Solids Ratio Comparisons

	COD/VS		*Organic N/VS*		*FOG/VS*	
	THS	*Product*	*THS*	*Product*	*THS*	*Product*
Average	1.48	1.38	0.065	0.055	0.045	0.010
Standard Deviation	0.15	0.16	0.014	0.010	0.024	0.004
Number	29	34	28	32	6	6

This data shows that the residual volatile solids in the VERTAD™ product contain constituents with lower COD per unit mass than the feed. This indicates that, on average over the entire test period, fats are differentially degraded in preference to combined carbohydrates and proteins. The same appears to be true for organic nitrogen (which correlates with protein) and fats/oils. This leads to the conclusion that carbohydrates are not as effectively degraded in the VERTAD™ process compared to fat and protein. The data also indicates significant potential variability in the mix of these constituents in the feed and the product. There is insufficient data to determine whether variation in the feed is passed through the process or whether process performance is also a factor in the variability in product mix character.

Another indication of the variability of the feed character occurred during Thanksgiving of 1999. Figure 2-10 shows the heat supplied to the reactor by the boiler over a 2 ½ day period beginning at midnight on Thanksgiving Day. There is no evidence that the concentration of feed solids to the reactor changed over this period, although it is possible. The figure shows a significant decrease in the heat supplied by the boiler beginning early Thursday morning. This continues through Friday evening and then begins to return to the initial energy addition condition. The exact reason for this situation is not known but possible causes include 1) an increase in the concentration of solids produced by the dissolved air flotation process or 2) an increase in the fat content of the feed resulting from preparation of Thanksgiving meals.

Figure 2-10
ENERGY INPUT
Thanksgiving Weekend 1999
Nov. 25-27



2.5.12.2 Supplemental Feed Study

On three occasions, during the latter part of the 1999 tests, supplemental feed in the form of vegetable oil was added to the reactor to observe process reactions. In one case sugar was also added. These tests were intended to determine the extent and duration of the process response to the addition of oil. Process responses being measured included heat release, temperature changes and oxygen transfer efficiency. Oxygen consumption was measured using an oxygen analyzer to monitor the reactor off gas. The first test consisted of two additions of oil. Ten pounds of soybean oil was pumped into the reactor (head tank) in a three-minute period. This quantity of oil, if fully degraded, would require 28.9 pounds of oxygen for conversion to CO_2 and H_2O. This conversion would release 167,000 BTU's of energy. The off gas O_2 concentration was recorded approximately every five minutes until it stopped decreasing. Then, an additional 25 pounds of soybean oil was pumped into the reactor (head tank) in a six-minute period. This additional quantity of oil, if fully degraded, would bring the required total demand from all added oil to 101 pounds of oxygen for conversion to CO_2 and H_2O. This conversion would release 585,000 BTU's of energy. The off gas O_2 concentration was recorded periodically until the evening. The oxygen analyzer was re-calibrated twice during the test readings. The adjustment in both cases was minor. The second test consisted of the addition of the same amount of oil as the first test, but all added as a single batch.

The third test involved the addition of 60 pounds of corn syrup and several bottles of high fructose drinks. This represents the addition of potential energy of 450,000 BTU's and an oxygen demand of 79 pounds. The syrup was added over a four-hour period. After four more hours without peaking the oxygen transfer efficiency, 20 pounds of oil was added. This provided an additional potential 325,000 BTU's of energy and 56 pounds of oxygen demand.

During all three tests the reactor was operating with diluted feed in order to maximize oxygen transfer efficiency. During the first test the reactor was operating at a 3.4 day HRT. During the last two tests the reactor was operating at a 5.4 day HRT. In none of the three cases was the oxygen demand measured through the entire period of response.

Figure 2-11 shows the observed differential oxygen uptake rate during this period (e.g. the difference between the initial oxygen transfer and the observed transfer). The response of the reactor to oil addition in the first two tests was rapid and dramatic. An analysis of the effect of the oil addition on heat added by the boiler is provided as a spreadsheet and graphic. This information indicates that oil addition continued to have an impact on heat addition for about 24 hours. The peak impact occurred after the O_2 transfer observation period. This may infer that we had even greater transfer efficiency than observed. This data also indicates that significant O_2 uptake was likely occurring for another 12 hours after observation. If uptake occurred at the average rate of 7 lbs per hour, the consumption of all of the added oil would be concluded. There is therefore evidence that most of the oil was degraded within 24 hours. Reactor temperatures increased by about 2°C following the oil addition.

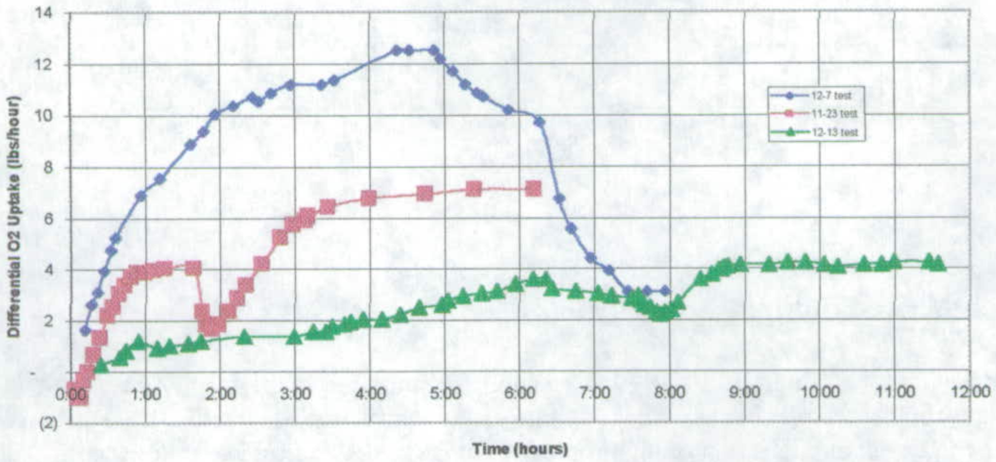
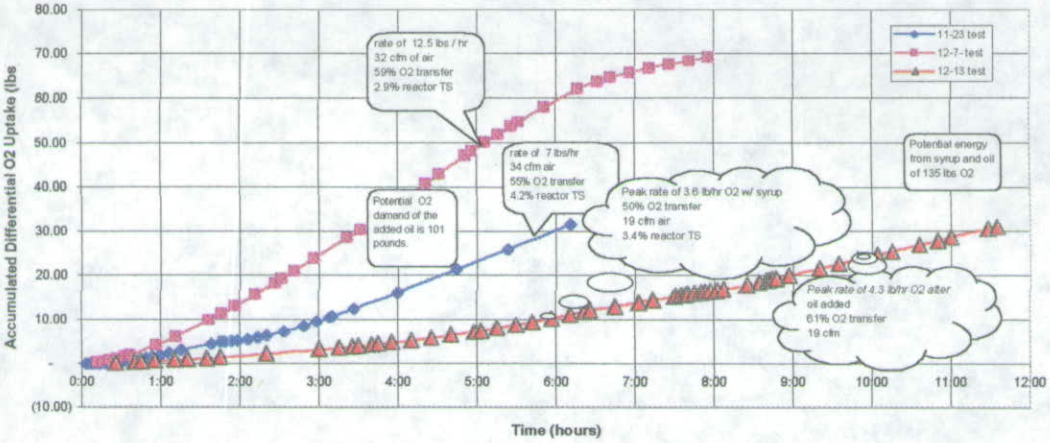


Figure 2-11: Supplemental O2 Uptake Rate
(for Oil Addition - Increase in O2 Uptake over background)

The addition of syrup in the third test created a much less dramatic response even though a similar amount of readily degradable, oxygen demanding organics was added. Figure 2-12 shows the accumulated increased oxygen transfer following supplemental feed addition.

Figure 2-12: Oil Addition Tests
Accumulative Increase in Oxygen Uptake over Background



During the period of observation only a fraction of the added oxygen demand was consumed in the reactor. Figure 2-13 shows the boiler energy input to the reactor during the test response. Although other factors can influence the boiler input, the response appears to show a reduction of about 400,000 BTU's over a 30-hour period. This reflects only somewhat more than the 325,000 BTU's added in the form of oil. The response was extended by limited oxygen availability after the addition of oil, which resulted from a low aeration rate. The reason for the apparent slow response and lack of energy release from the sugar addition is not known. This difference in response may indicate that even simple and highly degradable carbohydrates are processed differently (and less effectively) in the VERTAD™ process in comparison to fats.

Figure 2-13
ENERGY INPUT
Sugar and Oil Addition Test
Dec. 13, 1999

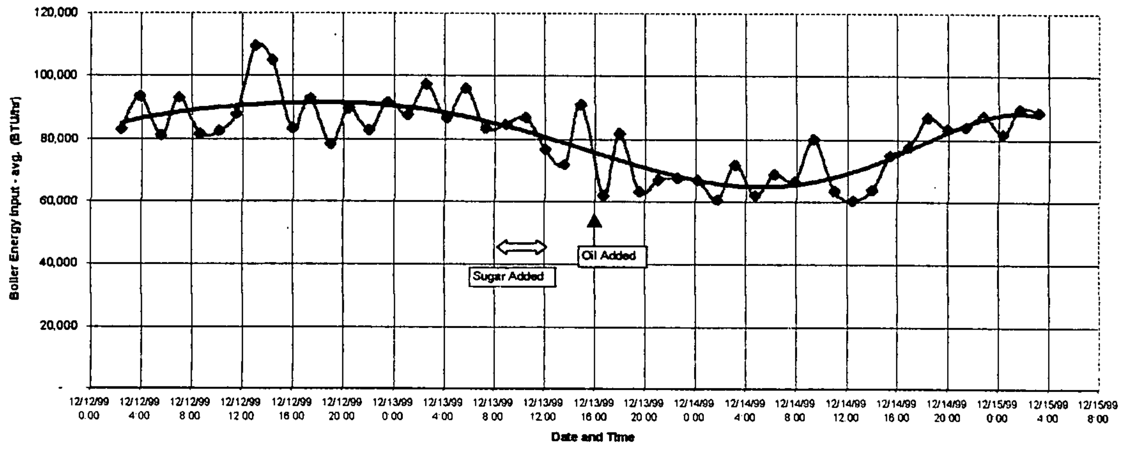
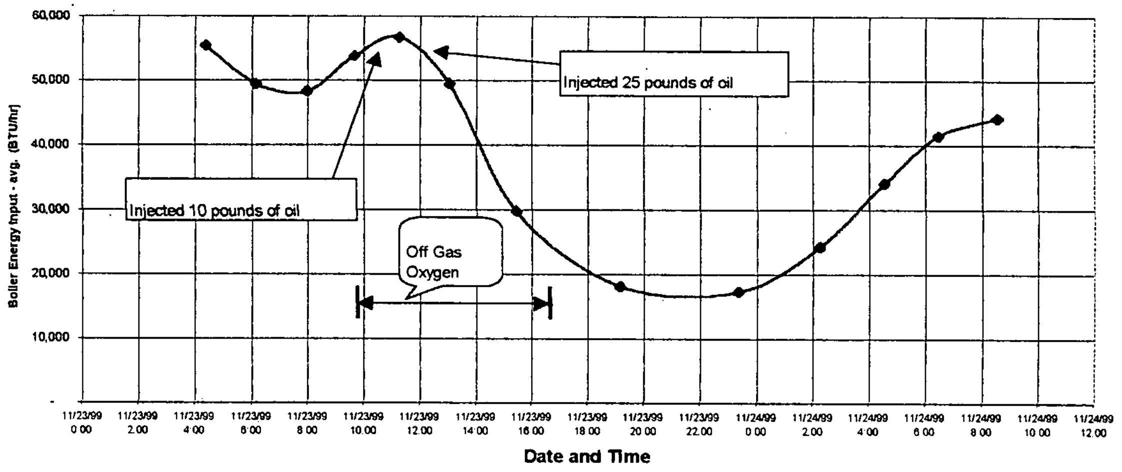


Figure 2-14 shows the response from the oil addition. The extent of energy input reduction and the shorter duration are in contrast to the sugar addition test. The extended response is at least partly due to limited oxygen transfer after oil addition. Sugar addition did not produce peak oxygen transfer.

Figure 2-14
Oil Addition Test 11-23-99



2.5.13 Energy Release and Loss

Heat Balance - A heat balance for Test 3 conditions is shown on Table 2-22. This table also provides estimated heat balances for detention times ranging from one to 10 days using assumed volatile solids reductions efficiencies and Test 3 operating conditions.

The heat balance estimates are derived from the previously described biological energy of degradation and heat loss to the environment. The estimated biological energy release is based on assumed volatile solids reductions for greater than four-day detention times. These assumed values are generally based on experience with ATAD systems. This information has not been developed for the VERTAD™ process. The reactor heat loss estimate includes any energy derived from aeration mixing. About 20 percent of the compressor energy is available as hot water (122F). Total possible recovery of the compressor energy is 95 percent. Most of this recoverable energy is available as rejected low-grade heat that can be used to heat buildings if the compressors are properly housed. Aeration energy losses are calculated based on psychrometric data for hot saturated air. The product heat loss is based on product mass and temperature measurements.

Based on these balances, the supplemental heat provided by the boiler to maintain the process operating temperature of 56°C during Test 3 was about $1.5 * 10^6$ BTU/day. Significant additional heat would be needed to operate at one through three day detention times (about $4*10^6$ BTU/day more for a one day detention time). Less energy is required to operate at five to ten day detention times. These energy demands are specific to the test facility.

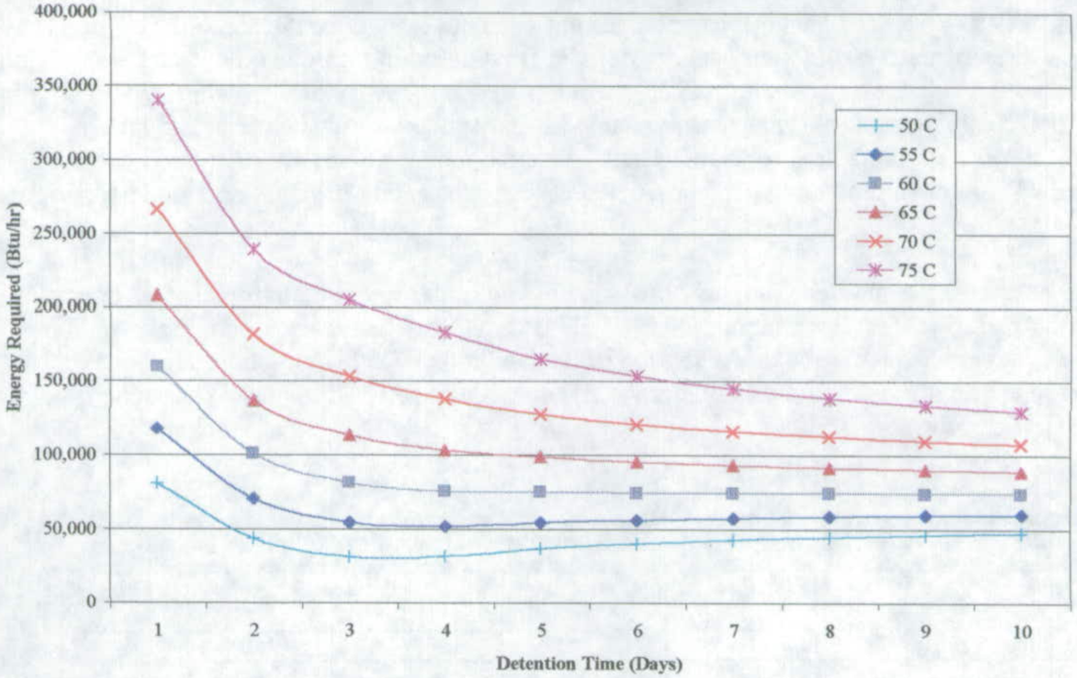
Table 2-22: VERTAD™ Heat Balance Summary - Operation @ 56°C

Detention Time		*Actual (4 day)*	*1 day*	*2 day*	*3 day*	*5 day*	*8 day*	*10 day*
Assumed VS Reduction Feed Solids	%	43	15	25	35	48	60	65
Supplemental Energy	10^3BTU/day	1,529	5,366	2,771	1,904	1,382	1,202	1,181
Aeration Energy	10^3BTU/day	-880	-1,226	-1,021	-953	-784	-613	-531
Biological Heat Addition	10^3BTU/day	2,191	3,052	2,544	2,374	1,953	1,526	1,322
Reactor Heat Loss to Environment	10^3BTU/day	1,390	1,390	1,390	1,390	1,390	1,390	1,390
Product Heat Loss	10^3BTU/day	1,450	5,802	2,901	1,934	1,160	725	580

Assumes 7,350 *BTU/lb.* VS destroyed and 0.7°C/hour heat loss

Heat balances for the demonstration reactor were also developed for operation at a range of temperatures and detention times as shown on Figure 2-15. The supplemental energy demand for maintaining a 70°C operating temperature at a 4 day HRT is estimated to be about three times the requirement for 56°C with Test 3 environmental and feed conditions.

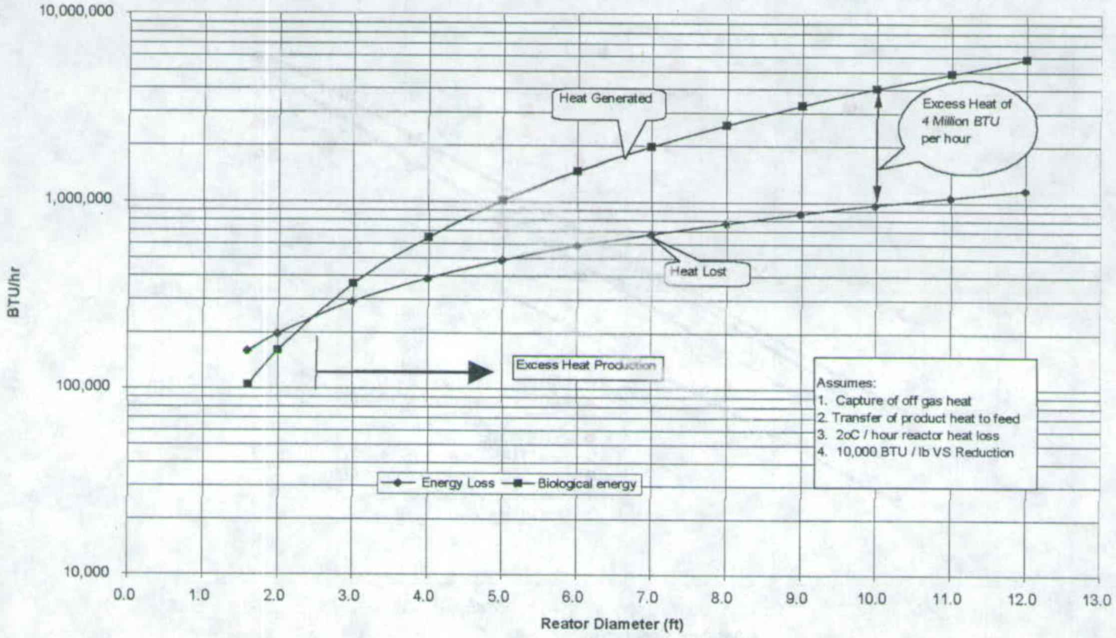
Figure 2-15: VERTAD Demonstration Facility Supplemental Energy Required for Thermophilic Operation



Process Scale Up - This thermophilic process relies on energy released during biological degradation to maintain operating temperatures. Therefore, the amount of energy available from degradation and the heat loss to the surroundings must be balanced to maintain desired operating temperatures. The test reactor has more heat loss than can be generated by degradation because of subsurface water and a high surface area to volume ratio. The test facility therefore requires a supplemental heat source. Supplemental heat will not be needed for larger reactors or reactors with insulation or more favorable subsurface geology.

Larger VERTAD™ reactors will have a very different heat balance due to lower surface area to volume ratio. In other words, less energy can be lost from the surface per volume of energy creating material in the reactor. Figure 2-16 compares the energy loss to the surroundings and off gases with the energy produced by the process biota. This indicates that for the same subsurface conditions experienced during Test 3, auto-thermal conditions would be attained with a single 2½ foot diameter reactor. Auto-thermal conditions could also be achieved with smaller reactors by insulating the reactor walls or locating in more favorable geological conditions. Clusters of small reactors would act like a larger reactor because the heat loss from each would be reduced by the proximity of the others.

Figure 2-16: VERTAD Reactor
Energy Production vs Energy Loss

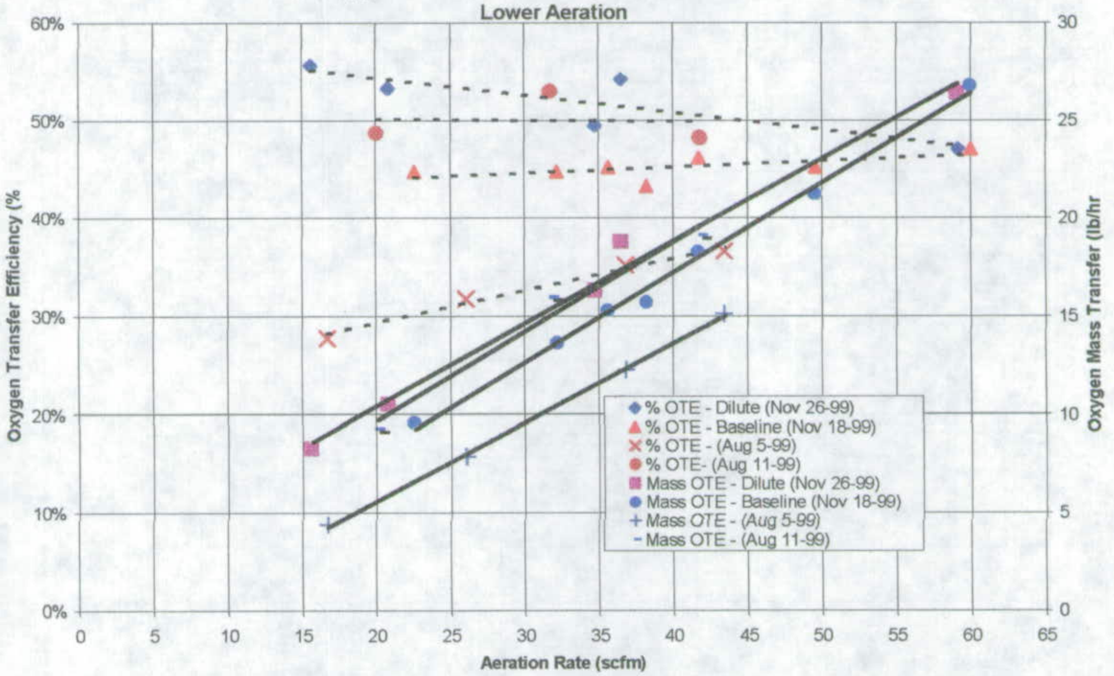


Reactors or groupings of reactors large enough to be auto-thermal would need to be provided with heat removal capability. Overheating of the reactor will result in a slowing of biological activity. The low-grade energy produced by reactor cooling may have potential value for use at the treatment facility or as an industrial heat source. For the test site conditions and third test operating conditions it is estimated that a 10 foot diameter reactor would produce about 48,000,000 BTU per day or 52,000 gallons per day of water at 70°C (158°F). An additional 16,000,000 BTU per day or 20,000 gallons per day of water at 55°C (131°F) could be recovered by cooling the product.

2.5.14 Aeration and Oxygen Transfer

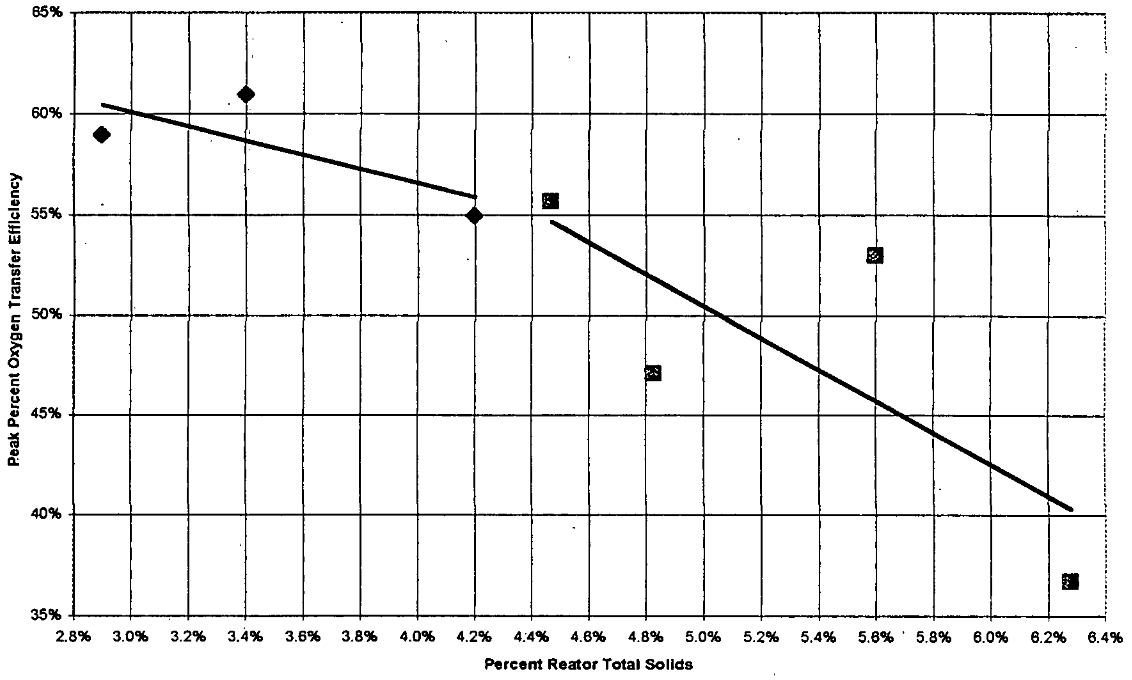
Head tank oxygen was measured for a variety of aeration rates and operating conditions using two oxygen gas analyzers. The results of some of this work was reported in the discussion of supplemental feed. With supplemental feed, oxygen transfer efficiencies (OTE) as high as 61 percent were measured. This is significantly above published results for surface tankage of 10 to 30 percent (Ref 12, pg. 497). Figure 2-17 shows the results of a series of tests conducted during the 1999 test period. These results were obtained during normal feed conditions. Testing was conducted to determine the best mix of upper and lower aeration. This testing documented that the most effective aeration method was to inject most of the air in the lower zone. The results presented in Figure 2-17 all involve primary aeration to the lower zone.

Figure 2-17
Oxygen Transfer - Efficiency and Mass Transfer
Lower Aeration



The results show considerable variation in OTE and mass oxygen transfer. The only clear trend is constantly increased mass transfer with increased aeration. The variability may be explained by Figure 2-18 which plots the peak measured OTE against the reactor solids content. This shows an apparent relationship with OTE potential decreasing as reactor solids content rises above 4.5 percent. A similar relationship was observed for VSR on Figure 2-5. The results indicate that a conservative design OTE for reactors operating with less than 4.5 percent solids content is 50 percent.

Figure 2-18: Peak Oxygen Transfer Viscosity Effects



The demand for oxygen associated with volatile solids reduction is also an important design criteria. Table 2-23 shows the measured oxygen demand per mass of volatile solids degraded as measured at four times during the 1999 test period. The data is for normal steady state operating conditions without influence of aeration setting changes or supplemental feed.

Table 2-23: Oxygen Uptake and Volatile Solids Ratio Comparisons

	HRT (days)	*VSR (%)*	*Oxygen Transfer*		
			Aeration (scfm)	*Transfer (pphr O_2)*	*O_2 Demand / lb VSR*
11/18/99	4	33	38	15.7	1.88
11/26/99	3.4	42	35	16.2	1.73
12/9/99	5.5	44	28	7.4	1.73
12/14/99	5.5	44	19	8.7	1.79

The results indicate an average oxygen demand of about 1.8 and a peak demand of about 1.9 pounds of oxygen per pound of volatile solids degraded. This result tends to support the concept of significant degradation of fat. However, the measured fat content of the feed has never been high enough to justify oxygen demand as high as shown on Table 2-23.

2.5.15 Operating Temperature

The demonstration facility has been successfully operated over a range of temperatures from 56°C to 67°C. Hotter temperatures appear to result in improved oxygen transfer and VSR. However, the amount of moisture removed with the off gas increases exponentially with temperature. A balance point appears to be 60°C for the design concept favored by the developer. This is based on 60°C being much better than 65°C in terms of water loss that has to be controlled with a biofilter. Ultimately, the larger the water loss, the bigger the heat exchanger on the biofilter. Designing for operation at 60°C is conservative (our testing indicates that there could be much greater destruction efficiency at 65-70°C, so long as the viscosity is controlled). Because of this conservatism, the plant potentially has additional capacity in case of permanent or temporary increased loading. The operating temperature of the reactor also influences the performance of flotation thickening. Somewhere above 60°C, ammonium bicarbonate is unstable and dissociates. This results in more ammonia in the off-gas (which has to be recaptured in a biofilter or scrubber) and less ammonium bicarbonate in solution for the acidification/float separation stage.

2.5.16 Reactor Mixing

Reactor mixing aeration characteristics have been observed while operating with water and with biological processing. The upper zone operates hydraulically as a plug flow reactor with large internal recycle and turbulent flow. This was confirmed before start-up by injecting acid into the recycle down-comer line and monitoring the pH and later using lithium as a tracer. In startup testing, the period between pH spikes (depending on the aeration rate) was 60 - 75 seconds corresponding to a riser velocity of approximately 1 m/sec (3.3 ft./sec) and down-comer velocity of 3.6 m/sec (12 ft./sec). Voidage was determined by measuring the liquid level change in the head tank where 3.3m (10.8 ft.) of voidage produced 0.3m (1 ft.) of head tank liquid level rise.

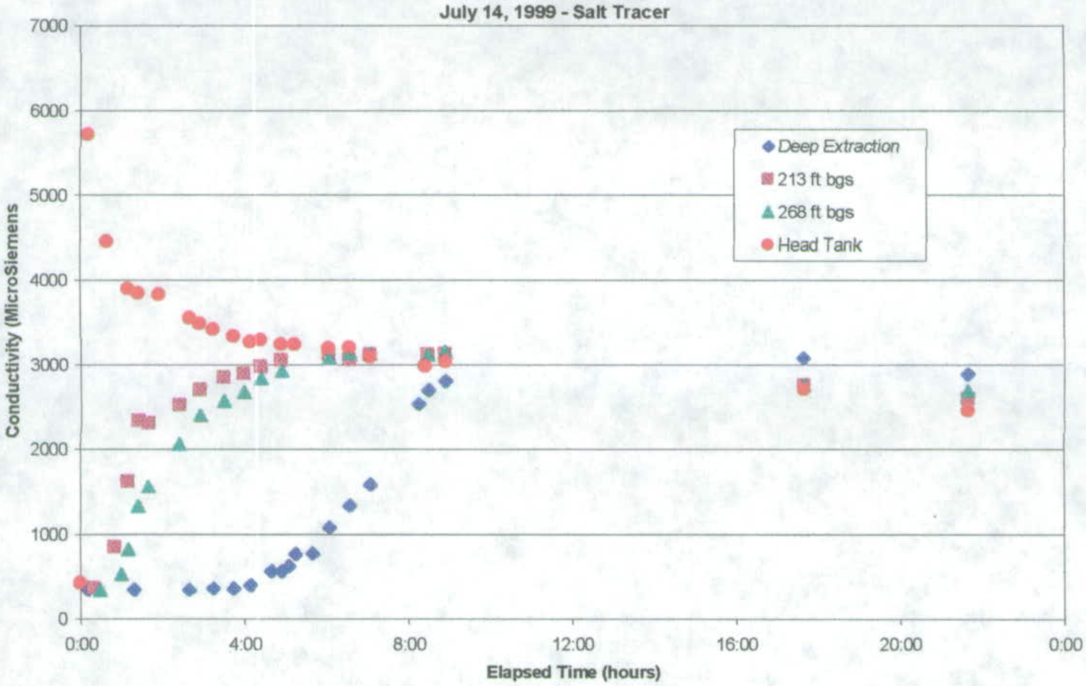
The lower reactor was designed to operate as a plug flow with localized back mixing. A lithium tracer study was conducted in clean water with a low rate of lower zone aeration (20 scfm). The lithium was pulse loaded. The results indicate that the lower zone is mixed over a period of 8 to 16 hours. The entire reactor behaves like two complete mixed reactors during this time period. Considering times longer than one day the reactor behaves like a single complete mixed reactor. Increased lower aeration rates may increase mixing in the lower zone and transfer between the upper and lower zones.

A lithium trace during biological operation indicated mixing throughout the aerated portion of the reactor. After injection of lithium to give 100ppb in the head tank, the first batch out contained only the background level of lithium (10 ppb) which indicates no mixing in the soak zone. The second batch (first from the lower aerated zone after lithium injection) contained 40 ppb and the third, fourth and fifth batches contained about 60 ppb. The test was done at aeration rates about 50 percent higher than the theoretical air requirement. There are about 11 batches in one reactor volume. Additional testing is desirable to identify the short duration mixing in the reactor. This is potentially critical for assuring that short-circuiting does not occur during Class A compliance.

The zone at the bottom of the reactor (soak zone) has no aeration or mixing. The oxygen is supplied entirely in the dissolved form during transfer from the lower reactor to the soak zone. Because there is limited circulation in the portion of the reactor above the soak zone, near saturation levels of dissolved oxygen (D.O.) can be achieved. In clean water, residual D.O. levels of 55 mg/L at 15°C were recorded. This residual D.O. was recorded at atmospheric pressure at the surface. It is estimated (from measurement made on other deep shaft plants) that the soak zone could achieve 75 mg/L at 15°C and 10 atmospheres of pressure. The lithium tracer study found no evidence of mixing in the lowest soak zone.

Before the initiation of the 1999 test series, a clean water mixing test was completed using sodium chloride. The results are shown on Figure 2-19. This test found complete mixing of the two mixed zones in about 4 hours. Evidence of salt movement through the soak zone was observed about 4 hours after salt addition. This test verifies the plug flow nature of the soak zone in the demonstration reactor. Similar verification or modeling would be necessary to document plug flow conditions in large reactors for purposes of Class A time and temperature compliance.

Figure 2-19
Vertical Mixing Study
July 14, 1999 - Salt Tracer



2.5.17 Off Gas and Odors

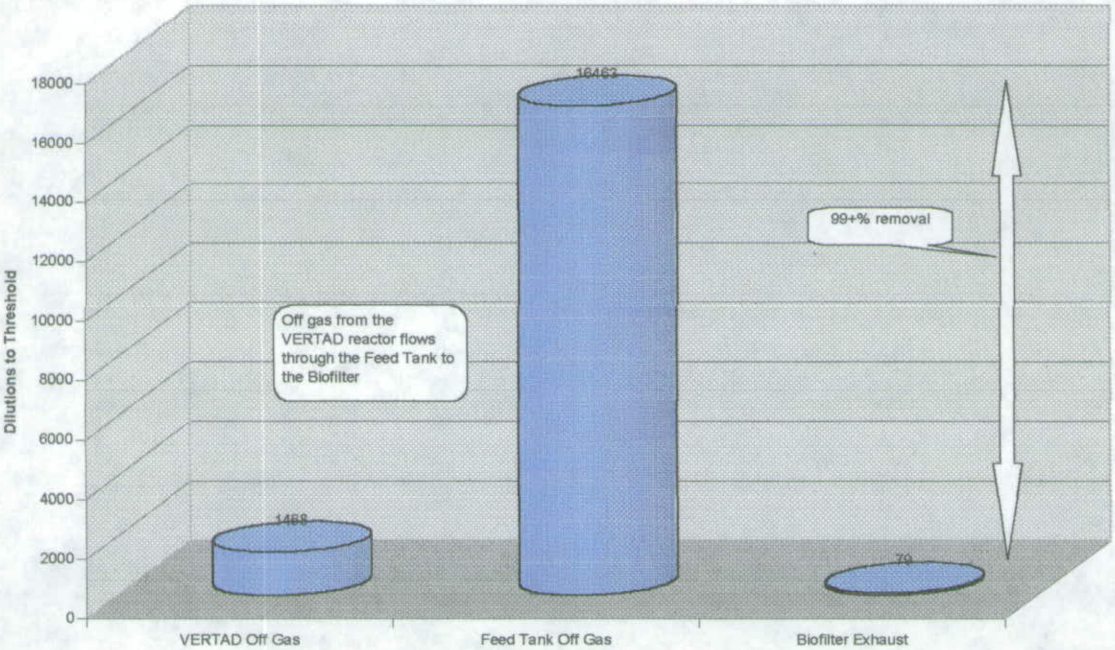
On December 20, 1999 samples were collected to measure the odors generated by the VERTAD™ process and the effectiveness of the biofilter for odor treatment. At the time of testing the compressor was operating at 41 percent (~36 scfm). The airflow to the biofilter (measured with a hot wire anemometer) was 670 scfm. Air temperature of biofilter feed flow was 20.5°C. The calculated loading on the biofilter was 2.8 feet per second face velocity. The nominal detention time was 1.4 minutes. The result of the odor panel work is provided as Figure 2-20. These results lead to the following conclusions:

1. Most of the VERTAD™ demonstration facility derived odor comes from the feed tank (16,463 D/T in 675 scfm) rather than the VERTAD™ reactor (1,468 D/T in 36 scfm).
   a) For the VERTAD™ process off gas prior to the feed tank, the odor panel used more odor character descriptions such as compost, earthy and vegetation although the typical sludge and manure are still used.

b) Passing the off gas through the feed tank changed the odor panel characterizations to focus on the sludge and manure type odors.

2. The biofilter removed 99.5 percent of the odor loading (16,463 D/T in and 79 D/T {avg} out).

Figure 2-20
Odor Panel Results



Off gas from the VERTAD reactor flows through the Feed Tank to the Biofilter

99+% removal

Demonstration and Evaluation
of
VERTAD™
Aerobic Thermophilic Digestion Process

Section 3

VERTAD™ Technology Implementation

Several options are available for incorporating the VERTAD™ technology into the King County wastewater management system. These options will be reviewed in the following section of the report. Most of these options were evaluated during the testing program. Those that were tested have been used in developing alternatives, including a conceptual design of facilities that could be used at both of the existing treatment facilities and at the planned new facility. Planning-level cost estimates are presented for these alternatives. The sensitivity of these estimates to changes in design assumptions is also evaluated.

3.1 Digestion Performance Objectives

The VERTAD™ process is designed to provide wastewater solids stabilization that would supplement or replace the anaerobic digestion process currently used by King County. In addition, VERTAD™ is capable of upgrading the resulting biosolids product from Class B to Class A pathogen control designation. Providing these functions at the lowest cost is always an overriding criteria for wastewater management. The objectives for VERTAD™ digestion alternatives relative to these basic criteria (stabilization, Class A and low cost) are explained in the following discussion.

3.1.1 Project Objectives

The VERTAD™ process was initially selected by the County for demonstration because it had the potential to assist with several goals for the West Point Treatment Plant:

- Reduced space requirement and the potential for future removal of existing anaerobic digesters
- Reduced truck traffic
- Improved odor control capability

3.1.2 Regulatory Issues

40 CFR 503 and WAC 173-308 define the Vector Attraction Reduction (VAR) and Pathogen control requirements for beneficial utilization of biosolids as practiced by King County. Although a variety of options are provided for satisfying these regulations the most common method of VAR compliance is by measuring the reduction in volatile solids content through the digestion process. A minimum of 38 percent reduction is required. Although the volatile solids test has the potential for inaccuracies it is the method utilized in this evaluation for documenting VAR compliance.

The regulations also provide several options for complying with the Class A pathogen control designation. The advantage of a Class A product is that it can be distributed to users without regulatory constraint or permits once the processing requirements are achieved. For VERTAD™, the compliance method of choice is Alternative 1 Regime D as defined in 40 CFR 503.32(a)(3)(ii)(D). This option allows compliance in a relatively short period over the typical operating temperature range. For example, at an operating temperature of 60°C, the required contact time for Class A compliance is five hours. At 65°C the contact time requirement drops to one hour.

3.1.3 Economic Factors

The economics of VERTAD™ are influenced by a variety of factors, most of which will be considered during the comparison of alternative costs:

Volume / Weight Reduction – Tests indicate that VERTAD™ performance results in less volume and weight of biosolids that must be transported and land applied or composted.
Dewaterability – The VERTAD™ process (including pH shift gas release flotation thickening) has the potential to deliver a thicker product while using less polymer.
Energy Recovery – VERTAD™ produces excess energy in the form of low grade hot water and hot air. These can be captured and used for heating solids prior to digestion and for space heating at the treatment facility and off site commercial development. Combining VERTAD™ with anaerobic digestion would provide the benefit of digester gas production as well as usable hot water. The optimum split between aerobic and anaerobic digestion has not yet been determined.
Energy Cost - Recent experience with energy price volatility demonstrates the importance of evaluating the impact of energy cost on the alternative comparisons.
Reactor insulation – Insulation of the VERTAD™ reactor would increase the recovery of heat and decrease the loss of heat to the groundwater. The cost benefits would need to be determined during design. Insulation of the reactor is not included in this evaluation of alternatives. The impact of heat flow into groundwater has also not been evaluated as a part of this study.
Nitrogen Content – VERTAD™ rejects a significant fraction of feed nitrogen as ammonia in the off gas and dewatering liquid side stream. This may result in a reduced fertilizer value for the product, but may also reduce application costs by allowing greater application rates per unit area.
Side Stream Characteristics – The impact of return liquid flow from the dewatering of VERTAD™ product on the wastewater treatment process has not been determined.
Class A Product – Permitting and oversight costs should be less for utilization of a Class A product compared to a Class B product. In addition, it may be possible to develop a product for local markets that would reduce current haul and application costs.

3.1.4 Aesthetic Issues

Odor – Odor control for the digestion process is included in the process design as described in detail later in this section. The odor of the product may also be an issue if the product is used for land application or as a feedstock for composting.
Product Aesthetic Characteristics – Since VERTAD™ is an aerobic process, the product is a lighter color and less odorous than the anaerobic product. This change will likely result in a need to redevelop existing markets to obtain acceptance or develop new markets for the material.

3.2 Potential Applications for Enhancement or Replacement of Existing Anaerobic Digestion Systems

The basis for funding of this demonstration program has been the promise of benefit to the County's wastewater management system. Several attractive options for utilization of this technology by the County are available and appear to be technically feasible based on the information gathered to date.

The VERTAD™ process performance indicates potential for several digestion processing configurations. The process can perform as a stand alone process, in sequence with anaerobic digestion (either mesophilic or thermophilic), or in combination with the Centridry process.

3.2.1 VERTAD™ Digestion

A VERTAD™ reactor can provide a stable, aerobic, Class A biosolids product. Dewaterability tests indicate that the VERTAD™ digested product dewaters very well in comparison to anaerobic solids particularly after pH shift gas release flotation thickening. Dewatering tests also support this approach with greater than 30 percent TS product from dewatering hot product. Initial tests indicate that the process can provide a Class A stabilized product on a very small processing footprint.

During demonstration testing, the VERTAD™ biological process was found to be very stable and required very little operator time or laboratory support for operations. The process was interrupted numerous times by non-process related mechanical and electrical problems, but quickly recovered with minimal operator involvement. This characteristic may indicate lower potential operational labor requirements for the process than anaerobic digestion.

3.2.2 Coupled VERTAD™ and Mesophilic Anaerobic Digestion

Linking the process with anaerobic digestion may provide substantial benefits including production of high quality biosolids product, high solids destruction efficiency, gas production, and good dewatering characteristics of the digested product. VERTAD™ could be used as an initial digestion step before anaerobic digestion or as a final step before dewatering. The VERTAD™ pretreatment step could be designed to maximize solids destruction (day SRT) or to solely meet Class A time-temperature requirements (1.5 day SRT). Following are some of the potential benefits of these combined processes:

VERTAD™ processing followed by anaerobic digestion operated for maximum solids destruction

a) Production of Class A biosolids at a three to four day retention time
b) Additional anaerobic digestion yields total volatile solids reduction of 65-70 percent
c) Hot product from VERTAD™ reduces heat requirements in following anaerobic process
d) Excess heat can be recovered for digester or general plant heating
e) pH shift gas flotation thickening prior to feeding anaerobic digestion minimizes the required size of the anaerobic tanks
f) Good mixing characteristics in anaerobic digester – no foam, scum
g) Good dewatering performance of final product

VERTAD™ operated for minimum time and temperature to produce a Class A product, followed by anaerobic digestion.

a) Class A time and temperature requirements achieved at 1.5 day SRT in VERTAD™; Required footprint is very small.
b) Hot, thickened solids delivered to anaerobic digestion minimizes digester volume and heating requirement
c) Estimated total volatile solids reduction in excess of 60 percent
d) Significant gas production
e) Good mixing characteristics in anaerobic digester

f) Good dewatering performance of final product

Anaerobic digestion followed by VERTAD™

a) Anaerobic digester operated to maximize methane production efficiency allowing reduction of digester retention time from about 30 days to 10 days or less
b) Small facility footprint
c) Class A Pathogen Control
d) Anaerobic pretreatment may improve solids destruction efficiency in VERTAD™
e) pH shift gas flotation thickening prior to dewatering may remove dissolved degradation products that inhibit dewatering, thereby improving dewatering performance and reducing polymer demand.
f) Aerobic product may reduce odor potential during dewatering, hauling and application.

3.2.3 Coupled VERTAD™ and Thermophilic Anaerobic Digestion

The excess heat generated by VERTAD™ is very compatible with maintenance of thermophilic temperatures in the linked anaerobic process. The advantage of this approach may be reduced retention requirements in the anaerobic digesters, which translate to less cost and even smaller footprint. VERTAD™ following anaerobic thermophilic digestion may provide a more aesthetically desirable and more easily dewatered product.

3.2.4 Coupled VERTAD™ and Centridry

Providing VERTAD™ digestion followed by Centridry would yield a small volume of Class A dried product. Composting would not then be required to provide a Class A product as is currently anticipated with anaerobic digestion followed by Centridry. Preceding drying with an aerobic process may also reduce the odor concerns currently associated with using Centridry with anaerobic sludges. Initial tests indicate that the drying performance of Centridry may also be improved as a result of the apparent greater dewaterability of the VERTAD™ product. Less polymer or drying heat may be required with the VERTAD™ product.

3.2.5 Coupled VERTAD™ and Short Duration Composting / Reduced Cost Class A Compost or Topsoil Product

Because VERTAD™ product satisfies Class A pathogen criteria, the composting process does not need to be designed to comply with Class A procedures. This provides potential cost savings by reducing the time required for processing. For example, GroCo, Inc., the County's current contractor may be able to reduce the current composting period from one year to one to six months. In addition, they may be able to reduce the amount of sawdust required as a bulking material because of the higher solids content. These changes could effectively increase the site capacity by ten fold while reducing the per ton processing costs. In a personal communication with Curley Winebrenner (Ref. 18) he speculated that a tip fee reduction of 25 to 30 percent may be justified. The product would likely have different characteristics than the current GroCo product. A significant marketing effort would be required to develop a market for this material that exceeds the current long term market for GroCo. There is also risk of losing a portion of the GroCo market as a result of a change in the product character.

Using a Class A biosolid as an ingredient in the manufacture of commercial topsoil products is a major potential market. This market is currently being explored by the County for the Centridry product, but may also be suitable for dewatered VERTAD™ product. It is very possible that between reduced haul costs for a local market and reduced monitoring and administrative costs, that the overall costs of biosolids management and reuse will be substantially reduced compared to current costs.

3.2.6 Use of VERTAD™ for Management of Food Waste and Biosolids

A recent evaluation of organics waste management methods by the County (Ref. 3) considered the potential for co-managing biosolids and source separated food waste and biosolids in a common facility. The VERTAD™ technology seems particularly well suited to this use. The ability of the process to rapidly degrade fats and protein with the production of heat and a Class A product may provide a cost effective method of managing source separated food waste. The potential for utilizing food waste management techniques seem to be particularly attractive as a feature of the planned North Treatment Plant.

3.2.7 Alternatives Selected for Comparison

Testing of the VERTAD™ process to date has focused primarily on documenting performance as a stand alone stabilization process. Bench scale evaluation of mesophilic anaerobic digestion following VERTAD™ was completed by University of Washington researchers (Appendix A). In addition, a single batch of VERTAD™ product was processed through the Centridry system.

Available facilities and resources prevented the testing of VERTAD™ following short duration anaerobic digestion. Any work with thermophilic anaerobic digestion was also prevented by lack of resources. So, although these process configurations hold significant promise, they will not be carried forward to the alternative comparison phase due to a lack of sufficient process design information. Similarly, the information available for processing VERTAD™ product through Centridry is insufficient to define design parameters.

Those process configurations that have sufficient design basis information for alternative development include:

1. VERTAD™ as a stand alone stabilization process
2. VERTAD™ followed by mesophilic anaerobic digestion.

This selection is based on availability of information and does not indicate that these process configurations are better in any way than the other configurations discussed above.

3.3 Full-Scale VERTAD™ Process Design Features

3.3.1 VERTAD™ Design Parameters Based on Demonstration Testing

Demonstration testing has identified several critical design parameters that determine the performance, sizing, and cost of VERTAD™ systems.

Reactor Viscosity – Oxygen transfer was found to be significantly reduced at reactor total solids content in excess of 4.5 to 5 percent TS. To allow high oxygen transfer rates the total solids content in the reactor has been limited to 4.3 percent for alternatives where 50 percent or greater oxygen transfer is desired. For alternatives with short detention time in the reactor, the design TS content was above 4.5 percent and lower oxygen transfer efficiencies were assumed.

Aeration Rate and Transfer Efficiency – The aeration rate is determined by oxygen transfer requirements. Mixing is adequate if aeration demands are met. Oxygen transfer efficiencies as high as 61 percent were measured during the demonstration testing. A maximum design value of 50 percent transfer has been selected for the purpose of developing the alternatives.

Flotation thickening - Initial testing indicates that the acidification of VERTAD™ product results in the release from solution of gas bubbles of carbon dioxide. These bubbles will float thicken the product in the same manner as a dissolved air flotation system. This was also shown to have a beneficial effect on dewatering. Research (Ref. 7) has found that biopolymers and sodium associated with the liquid fraction of digested solids reduce dewatering effectiveness and increase polymer demand. The flotation step may separate these undesirable constituents from the solids prior to the dewatering process, thereby reducing polymer demand.

In alternatives involving dual digestion, a critical factor in the design was the internal concentration in the anaerobic digester. Since there is little data regarding operating a digester at high solids concentration, this parameter was conservatively set at 4.5 percent. So despite the fact that the concentration of the float thickened solids can be up to 12 percent, significantly reducing downstream tankage and dewatering equipment, approximately 7 percent was the maximum allowable float concentration to stay within the digester constraints.

In straight VERTAD™ options, a 10 percent float thickness was used. A 95 percent capture efficiency was used for the flotation cell. The acid requirement to float is 0.0004 gallons of 93 percent H_2SO_4 per gallons of VERTAD™ product. Surface area of the flotation cells is set by the allowable surface loading rate of 1.8 pounds of solids per square foot per hour.

Linked Anaerobic Digestion – The University of Washington evaluation found that anaerobic digestion detention times of 11 and 15 days preceded by four days of VERTAD™ digestion provided a high degree of VS reduction (63 and 66 percent). Since the design detention times for anaerobic digestion are significantly higher than those used in the study, the assumed VS reduction used in the evaluation are 65 percent for Class A VERTAD™ and 70 percent for four days in VERTAD™ followed by anaerobic digestion.

Temperature and pH – Somewhere above 60°C, ammonium bicarbonate is unstable and decomposes. This results in more ammonia in the off-gas (which has to be recaptured in a biofilter or liquid scrubber) and less ammonium bicarbonate in solution for the acidification/float separation stage that follows the VERTAD™. The ammonia and CO_2 are liberated from this decomposition. The ammonia re-adsorbs into solution much more readily than the CO_2, raising the pH from approximately 8 to 8.8. In terms of design, this means an increased sulfuric acid requirement (in reactors operating above 60°C) in the flotation stage that follows to depress the pH to the point where CO_2 is released (approximately pH 5).

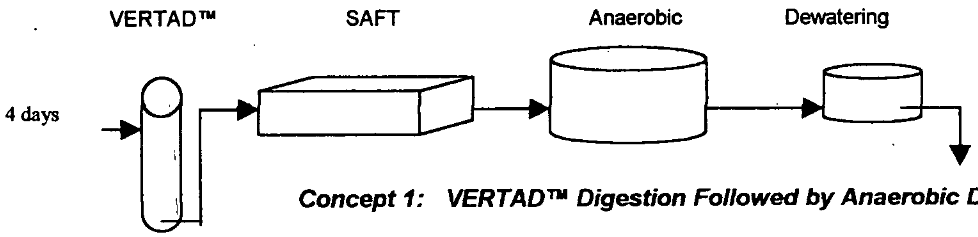
In addition, a 60°C operating temperature is much better than 65°C in terms of water loss from the reactor that has to be managed with a biofilter. Ultimately, the larger the water loss, the bigger the heat exchanger on the biofilter. Designing for operation at 60°C is conservative (testing indicates that there could be much greater destruction efficiency at 65-70°C, so long as the viscosity is controlled). Because of this conservatism, the plant has additional capacity to handle peak loading conditions.

Residence Time – Volatile solids reduction of 38 to 45 percent have been demonstrated when operating VERTAD™ at three and four day detention times. For purposes of alternative development the conservative detention time of four days will be used.

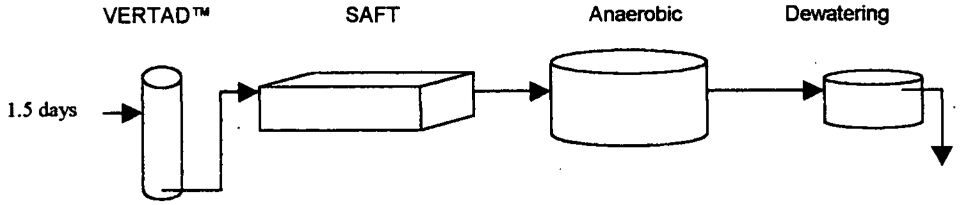
3.3.2 VERTAD™ Process Design Concepts

Three process configurations using VERTAD™ are evaluated in this analysis. The VERTAD™ process design conceptual flow diagrams for these process configurations are provided on Figure 3-1.

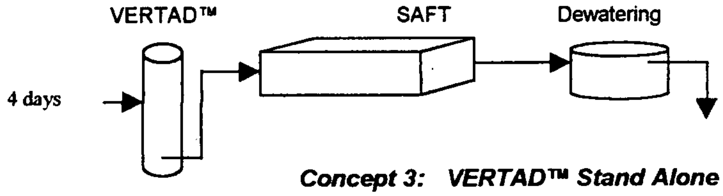
Figure 3-1: VERTAD™ Process Flow Diagrams



Concept 1: VERTAD™ Digestion Followed by Anaerobic Digestion



Concept 2: Class A VERTAD™ Followed by Anaerobic Digestion



Concept 3: VERTAD™ Stand Alone Digestion Process

3.3.2.1 Concept 1: VERTAD™ Digestion Followed by Mesophilic Anaerobic Digestion

This concept is designed to take advantage of the known optimum capabilities of the aerobic and anaerobic processes. VERTAD™ can theoretically be effective for pre-processing solids before anaerobic digestion or for providing the final stabilization following an initial anaerobic digestion. In either case, the detention times used are expected to be shorter than normally used by either process alone. Gas production, energy requirements and product quality would be expected to vary significantly between these two approaches. Since only the approach with VERTAD™ preceding anaerobic digestion has been tested to date, this is the alternative that will be evaluated.

The objective of this alternative is to provide sufficient aerobic thermophilic and mesophilic anaerobic digestion capacity to produce a well stabilized, Class A biosolid that dewaters to a high solids content with relatively low polymer demand. To accomplish this the selected detention times are four days in the VERTAD™ reactor, which provides a Class A, stabilized product. It is then followed with 24 day detention mesophilic anaerobic digestion to provide stabilization and compliance with VAR. The 24 day detention was based on the University of Washington study which found that performance is excellent with a 15 day detention time. The 24-day average annual detention provides about 16 days detention time at peak loading conditions.

The VERTAD™ process consists of a deep aerated reactor with high and low rate mixing zones and an unmixed Class A contact zone. Flow through the reactor is from top to bottom. The process off gas is treated by a liquid media biofilter located in the reactor head tank. Since large volume VERTAD™ reactors produce excess heat, they must be cooled to maintain the desired operating temperatures for optimum performance. Primary cooling is provided by a sludge to water heat exchanger that draws from the reactor, cools, and returns the material to the reactor. Heat is also removed from the biofilter to maintain appropriate temperatures for the micro-organisms that degrade the captured odorous compounds. A water to water heat exchanger is used for this purpose. The product from the reactor is used to pre-heat the feed solids in a sludge to sludge heat exchanger.

3.3.2.2 Concept 2: VERTAD™ for Class A followed by Mesophilic Anaerobic Digestion for VAR

This concept is based on the finding of the demonstration testing that the VERTAD™ digestion process can degrade a significant portion of the volatile fraction during a two-day detention time. The objective of this process design is to use VERTAD™ to satisfy Class A pathogen requirements with a minimum detention time and provide the balance of stabilization with mesophilic anaerobic digestion. Based on energy balance information developed during the demonstration study a 1.5-day detention time was selected as the threshold of auto-thermal operation of the VERTAD™ reactor. Sufficient energy is generated in the VERTAD™ process to heat both the aerobic and the anaerobic processes.

This system has an odor control and heat management system similar to the previous concept except that a heat exchanger is not provided to cool the reactor. This is not needed because the reactor is designed to operate at near energy balance. The other two heat exchangers can be operated to maintain appropriate temperatures in the reactor.

3.3.2.3 Concept 3: VERTAD™ Digestion

VERTAD™ can serve as the sole stabilization process. Demonstration testing indicates that thermophilic aerobic digestion can be completed with a solids retention time of four days. Digestion is not as extensive as provided by combined anaerobic / aerobic systems, so product dry weight quantities are greater than for the other two concepts. Demonstration testing found that the VERTAD™ product can be floated by adding acid to release dissolved carbon dioxide as gas. This separation provides a thicker feed to dewatering and also improves dewatering performance relative to solids content and polymer demand. This flotation process has been included in all of the VERTAD™ process flow sheets.

3.3.3 Comparative Anaerobic Digestion Processes

In addition to costs for these VERTAD™ alternatives, costs are also presented for comparable anaerobic digestion based alternatives. The costs for these alternatives are based on cost estimates developed by Brown and Caldwell for the South Treatment Plant (Ref. 2). These alternatives include the following design assumptions:

Mesophilic Anaerobic Digestion – This is the existing digestion technology used at the South and West Point Treatment Plants. Generally the process provides at least 55 percent volatile solids reduction with a 24 day detention time. The product is assumed to dewater to 20 to 24 percent total solids (with centrifugal dewatering). Gas production is typically 15 cubic feet per pound of volatile solids removed.

Thermophilic-Mesophilic Digestion – This technology uses a shorter duration thermophilic digestion step followed by a longer mesophilic step. Thermo-meso digestion has been operational at full-scale by other facilities for several years. A pilot-scale demonstration of the technology has been completed at the West Point Treatment Plant. Generally the process provides 65 to 70 percent volatile solids reduction with a 20-day detention time. The product is predicted to dewater to 23 - 27 percent total solids (with centrifugal dewatering). Gas production is typically 15 cubic feet per pound of volatile solids removed. The plan for the South Plant consists of converting two of the four existing digesters to be able to operate as thermophilic digesters. One digester at a time would operate thermophilically while the three remaining would operate mesophilically, in parallel.

Thermo-meso Digestion with Batch Time-Temperature Class A Process – This process is similar to the previous except a batch time-temperature process is added between the thermophilic and mesophilic phases in order to achieve compliance with Class A criteria in a manner similar to the soak zone of the VERTAD™ reactor.

Table 3-1 provides the equipment and facility design sizing criteria for each of these concepts.

Table 3-1: Equipment and Facilities Design Sizing Criteria

Category	*Units*	*Sizing Criteria*
VERTAD™ Reactor Depth	feet	350
VERTAD™ Reactor Detention Times		
VERTAD™ to Anaerobic-Concept 1	days	4 (at annual average flow)
Class A VERTAD™ to Anaerobic-Concept 2	days	1.5 (at annual average flow)
VERTAD™ Alone-Concept 3	days	4 (at peak 3 week flow)
Anaerobic Digester Detention Times		(all at average annual flow)
Mesophilic Anaerobic	days	31
Thermo-Meso Digestion	days	6.2/18.5
Thermo-Meso-hot tank in Series	days	6.2/18.5/0.5
VERTAD™ to anaerobic	days	24
Class A VERTAD™ to anaerobic	days	24 to 28
Overall Volatile Solids Reduction		
VERTAD™ Alone	%	40
Mesophilic Anaerobic	%	55
Thermo-Meso Digestion	%	65
Thermo-Meso-hot tank in Series	%	65
VERTAD™ to anaerobic	%	70
Class A VERTAD™ to anaerobic	%	65
VERTAD™ Class A Compliance		
Time	hours	10 (twice required time)
Temperature	°C	60
VERTAD™ oxygen Transfer Efficiency		
4 day VERTAD™	%	50
1.5 day Class A VERTAD™	%	35 to 40
Polymer Demand		
VERTAD™ Alone	lb/DT	20
Mesophilic Anaerobic	lb/DT	35
Thermo-Meso Digestion	lb/DT	35
Thermo-Meso-hot tank in Series	lb/DT	35
VERTAD™ to anaerobic	lb/DT	25
Class A VERTAD™ to anaerobic	lb/DT	25
VERTAD™ATD Gas Flotation Product	%TS	6.9 to 10
Dewatering Performance		
VERTAD™ Alone	%TS	30
Mesophilic Anaerobic	%TS	23.4
Thermo-Meso Digestion	%TS	25
Thermo-Meso-hot tank in Series	%TS	25
VERTAD™ to anaerobic	%TS	30
Class A VERTAD™ to anaerobic	%TS	30

3.3.4 Operational Design Features

The vertical, deep reactor configuration of the VERTAD™ reactor raises several specific design issues. For purposes of this cost estimate it is assumed that the following features are included in the VERTAD™ facilities being evaluated:

1. *Materials and Corrosion* – Standard mild steel casing material has been shown to provide long lasting service and corrosion resistance without special coatings and is the assumed primary material for use throughout the reactor.
2. *Down Hole Equipment Repair* – All process piping and equipment that is installed in the reactor will be designed for easy removal from the reactor as required for repair. Surface access hatches are assumed for removal with readily available rental equipment.
3. *Struvite Management* – Again, all internal equipment will be removable for maintenance if struvite deposits accumulate.
4. *Grit Accumulation Removal* – Experience indicates that grit will be removed with product and will not accumulate in the bottom of the reactor. However, the access will allow removal of any accumulation using small diameter piping and an air lift.

3.3.5 Alternative Descriptions

Several alternatives for VERTAD™ implementation were developed for the West Point and South Treatment Plants and the planned North Treatment Plant. A planning period of 20 years was used to establish design criteria. In general, VERTAD™ facility would provide a large amount of digestion capacity in a small amount of space. Construction would require few modifications to existing equipment and would not disrupt operations. Implementation could be staged based on projected loads. There is the potential to nearly double the solids digestion capacity at the existing plants in a very small footprint without significantly modifying existing facilities.

Table 3-2 provides a summary of design loading data for the treatment plant alternatives. The data is based on projected year 2019 solids loading.

Table 3-2: Treatment Plant Alternatives – Design Data

Category	*Units*	*Design Data*
Solids Loading (Average Annual)		
West Point Plant	DT/day	110
South Plant	DT/day	103
North Plant	DT/day	34
Solids Content (Average Annual)		
West Point Plant	%TS	5.5
South Plant	%TS	6.25
North Plant	%TS	5.5
Flow (Average Annual)		
West Point Plant	gpd	480,000
South Plant	gpd	395,000
North Plant	gpd	150,000

3.3.5.1 West Point Treatment Plant

The West Point Treatment Plant has adequate digestion capacity (with approved expansion) to serve the facility for many years. This analysis is intended to document the potential for removal of digesters as indicated in the initial objectives for the VERTAD™ demonstration project. Both alternatives being evaluated involve providing initial VERTAD™ processing of the solids prior to anaerobic digestion in existing anaerobic digesters. The VERTAD™ capacity allows production of a Class A product using three or four of the existing anaerobic digesters. The balance of the existing digesters could be mothballed for later use or removed.

Alternative WP-1 utilizes concept 1 as previously described. The primary design features of this alternative include:

- Four – 12.3 foot diameter by 350 feet deep reactors to provide four days of treatment at average annual solids loading.
- Compressors to provide 8,000 scfm of air pressurized to 150 psi.
- Four sulfuric acid flotation units to thicken the VERTAD™ product prior to anaerobic digestion.
- Three operational anaerobic digesters providing 24 days of treatment.

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration
- Class A biosolids product
- Additional excess digester gas beyond that needed for digester heating

Appendix B provides detailed process flow diagrams for all alternatives discussed in this section.

Appendix C includes the detailed VERTAD™ facility design basis loading and sizing calculation spreadsheets.

Alternative WP-2 utilizes concept 2 as previously described. The primary design features of this alternative include:

- Two – 11 foot diameter by 350 feet deep reactors to provide 1.5 days of treatment at average annual loading
- Compressors to provide 3,600 scfm of air pressurized to 150 psi.
- Four sulfuric acid flotation units to thicken the VERTAD™ product prior to anaerobic digestion
- Four operational anaerobic digesters providing 27 days of treatment

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration
- Class A biosolids product
- Additional excess digester gas beyond that needed for digester heating

3.3.5.2 South Treatment Plant

Several alternatives for adding digestion capacity to the South Treatment Plant are currently under consideration (Ref. 3). The anaerobic digestion based alternatives that will be compared to VERTAD™ alternatives for this facility include:

- Thermophilic-Mesophilic Anaerobic Digestion
- Thermo-Meso digestion with Class A Holding tanks

The VERTAD™ alternatives being evaluated include all three concepts defined previously.

Alternative SP-1 utilizes concept 1 as previously described. The primary design features of this alternative include:

- Three – 12.8 foot diameter by 350 feet deep reactors to provide four days of treatment at average annual loading
- Compressors to provide 7,300 scfm of air pressurized to 150 psi
- Three sulfuric acid flotation units to thicken the VERTAD™ product prior to anaerobic digestion
- Two operational anaerobic digesters providing 23 days of treatment

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration
- Class A biosolids product
- Additional excess digester gas beyond that needed for digester heating

Alternative SP-2 utilizes concept 2 as previously described. The primary design features of this alternative include:

- Two – 9.6 foot diameter by 350 feet deep reactors to provide 1.5 days of treatment at average annual loading
- Compressors to provide 3,600 scfm of air pressurized to 150 psi
- Four sulfuric acid flotation units to thicken the VERTAD™ product prior to anaerobic digestion
- Three operational anaerobic digesters providing 27 days of treatment

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration
- Class A biosolids product
- Additional excess digester gas beyond that needed for digester heating

Alternative SP-3 utilizes concept 3 as previously described. The primary design features of this alternative include:

- Five – 12.3 foot diameter by 350 feet deep reactors to provide six days of treatment at average annual loading and four days of treatment at peak three week loading
- Compressors to provide 12,000 scfm of air pressurized to 150 psi
- Five sulfuric acid flotation units to thicken the VERTAD™ product prior to dewatering
- No operational anaerobic digesters

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration
- Class A biosolids product

3.3.5.3 North Treatment Plant

The North Treatment Plant is a planned new facility to provide treatment capacity for the northern portion of the County service area. Several alternatives for providing digestion capacity to the North Treatment

Plant have potential. Anaerobic digestion based alternatives that will be compared to VERTAD™ alternatives for this facility include:

- Mesophilic Anaerobic Digestion (Base case)
- Thermo-meso digestion with Class A holding tanks

The VERTAD™ alternatives being evaluated include all three concepts defined previously.

Alternative NP-1 utilizes concept 1 as previously described. The primary design features of this alternative include:

- Two – 10 foot diameter by 350 feet deep reactors to provide four days of treatment at average annual loading
- Compressors to provide 2,400 scfm of air pressurized to 150 psi
- Two sulfuric acid flotation units to thicken the VERTAD™ product prior to anaerobic digestion
- Two operational 67 foot diameter anaerobic digesters providing 24 days of treatment

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration
- Class A biosolids product
- Hot water and digester gas beyond that needed for digester heating

Alternative NP-2 utilizes concept 2 as previously described. The primary design features of this alternative include:

- Two – 6.7 foot diameter by 350 feet deep reactors to provide 1.5 days of treatment at average annual loading
- Compressors to provide 1,200 scfm of air pressurized to 150 psi
- Two sulfuric acid flotation units to thicken the VERTAD™ product prior to anaerobic digestion
- Two operational 73 foot diameter anaerobic digesters providing 24 days of treatment

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration
- Class A biosolids product
- Hot water and digester gas beyond that needed for digester heating

Alternative NP-3 utilizes concept 3 as previously described. The primary design features of this alternative include:

- Two – 12 foot diameter by 350 feet deep reactors to provide six days of treatment at average annual loading and 4 days of treatment at peak three week loading
- Compressors to provide 3,600 scfm of air pressurized to 150 psi
- Two sulfuric acid flotation units to thicken the VERTAD™ product prior to dewatering
- No operational anaerobic digesters

This alternative is projected to produce:

- Dewatered solids at 30 percent total solids concentration

- Class A biosolids product
- Hot water

3.4 Implementation Costs Comparison

3.4.1 Basis of Cost Estimates

3.4.1.1 Objectives:

The objectives for the cost estimate include:

1. Develop cost estimates that are typical and consistent with estimates that the County has used for previous evaluations.
2. Provide a cost estimate that provides a high level of confidence for the County that the costs are reliable and realistic.
   a. Support costs for components that are unusual (in the County's experience) with documented construction cost information from other facilities (Appendix C).
   b. Use cost experience from County facilities when available.
3. Provide a cost protocol that can be used to easily update the estimates based on the results of further testing.

3.4.1.2 Cost Estimating Procedure

The procedure for estimating alternative costs for VERTAD™ implementation uses an established cost estimate format developed by Brown and Caldwell for the South Treatment plant digestion evaluation (Ref. 3). Capital costs are based on design and equipment sizing information provided by NORAM Engineering and Constructors, Ltd. which hold the patent for the VERTAD™ process. NORAM has provided a detailed schedule of services and equipment that would be provided by NORAM as a service associated with the County use of the technology. NORAM also provided a guaranteed price for the services and equipment that they would provide as well as a cost estimate for the balance of facilities required for a fully functional VERTAD™ plant. In general NORAM will provide the following:

- Process and Instrumentation drawings for VERTAD™ and associated features including odor control
- Complete layout and configuration drawings, including detailed piping schematics
- Electrical line drawings
- All mechanical equipment and instrumentation
- Coordination and review of detailed structural and electrical design drawings and specifications
- Construction oversight for NORAM provided equipment and systems
- Commissioning and start up services

Reactor drilling and installation costs are based on experience at 11 North American installations of similar wastewater reactors (Appendix C). Larger reactor costs are based on mining cost proposals developed by a Contractor.

Contractor overhead and profit (OH&P) was adjusted based on the OH&P included in each source cost estimate in order to provide a consistent total multiplier for all cost estimates. For example, the NORAM cost estimates already included an OH&P factor. The OH&P multiplier was therefore adjusted to provide the same total as the B&C cost estimates. Contingencies were selected based on the reliability of the cost source. For example, the guaranteed cost quote from NORAM was considered to be a very reliable cost and was adjusted with a small contingency factor. Other cost sources received a relatively high adjustment typical for planning level cost estimates.

As a check on the NORAM estimate, an independent cost analysis was completed for alternative NP-2. This analysis agreed with the NORAM costs to within 5 percent.

Operation and Maintenance costs are based on information included in the spreadsheet cost program develop for Ref 3. Consistent methods were used to estimate the costs and revenues associated with labor, power, gas generation and utilization, polymer, and biosolids management.

The annualized and present worth costs are calculated for a 17 year return period to remain consistent with the Brown and Caldwell evaluation.

3.4.2 West Point Treatment Plant

For the West Point Plant, only the capital costs of the VERTAD™ alternatives are compared. Operating and maintenance costs for current operations and other alternative technologies were not evaluated. (Present worth analyses are presented for the South and North Treatment Plants.). The primary difference between the two alternatives considered is that only three anaerobic digesters would need to be operated along with VERTAD for Alternative WP-1 while four anaerobic digesters would be required for Alternative WP-2.

3.4.2.1 Capital Cost Estimates

Table 3-3 provides the capital cost estimates for the West Point Treatment Plant alternatives. The minimum detention time VERTAD™ option (WP-2) has a significantly lower capital cost since only two reactors are required, compared to four for the full digestion VERTAD™ option (WP-1). Since the existing anaerobic digesters have already been constructed, those sunk costs are not considered in the cost comparison.

Table 3-3: West Point Treatment Plant Alternatives - Capital Cost Estimate (All Costs in $1,000's)

Category	*Alt WP1*	*Alt WP2*
DIGESTION		
Site Work	$50	$50
Structures	1,390	930
Equipment & Mech.	1,160	760
Electrical/I&C	340	270
Testing/Start-up	Incl.	Incl.
NORAM provided equipment and engineering	7,010	4,130
Cased Reactors	6,640	3,660
Subtotal	16,590	9,800
Contractor Indirects, OH&P 35%	17	17
NORAM estimate 11.6%	1,105	652
NORAM provided 5%	350	207
Subtotal	18,063	10,676
Contingency 30%	1,442	943
Drilling 10%	722	399
NORAM provided 5%	382	225
Subtotal	20,610	12,243
Sales tax 8.4%	1,731	1,028
Subtotal	22,341	13,271
Allied Cost 35%	7,819	4,645
Less NORAM engineering 15%	(1,173)	(697)
Total Capital Expenditure	***$28,987***	***$17,219***

3.4.2.2 Operation and Maintenance Costs

The anticipated impacts on O&M costs of VERTAD™ implementation at West Point include the following:

- One or two digesters would be taken out of service resulting in O&M savings.
- Heat generated in VERTAD™ would replace or supplement existing boiler system.
- Power demand would be higher due to aeration energy requirement.
- Better dewaterability of digested product would yield higher cake solids and lower polymer demand.
- Fewer wet tons of biosolids would be hauled due to higher solids destruction efficiency and better dewatering.

The impacts on O&M costs of the proposed facilities at West Point would be similar to those at South Plant, which are discussed below.

3.4.3 South Treatment Plant

3.4.3.1 Capital Cost Estimates

Table 3-4 provides the capital cost estimates for the South Treatment Plant alternatives. The highest capital cost is associated with Alternative SP-3 since five VERTAD™ reactors are required to handle the peak three week solids loading rate. Alt SP-2 (minimum detention time VERTAD™) has a similar capital cost as the Class A – thermo-meso digestion alternative. The thermo-meso digestion alternative has the lowest capital cost because it can be accomplished by modifying the existing digesters. The thermo-meso digestion – Class A alternative requires four batch time-temperature hold tanks, otherwise using existing tankage.

The capital cost for dewatering facilities is based on Brown and Caldwell's estimate for conversion from belt filter presses to centrifuge dewatering. More centrifuge capacity is assumed required for alternative SP-3 which yields more flow and solids for dewatering.

Table 3-4: South Treatment Plant Alternatives - Capital Cost Estimate (All Costs in $1,000's)

Category	***TMD****	***Class A***	***Alt SP1***	***Alt SP2***	***Alt SP3***
DIGESTION					
Site Work	$140	$140	$606	$606	$606
Structures	1,820	3,300	1,330	890	1,940
Equipment & Mech.	2,270	3,310	1,060	720	1,400
Electrical/I&C	460	660	320	270	390
Testing/Start-up	110	120	incl.	incl.	incl.
NORAM provided equipment and engineering			6,290	3,900	8,470
Cased Reactors			6,140	3,140	7,000
Subtotal	4,800	7,530	15,746	9,526	19,806
Contractor Indirects, OH&P 35%	1,680	2,636			
NORAM estimate 11.6%			1,097	653	1,315
NORAM provided 5%			315	195	424
Subtotal	6,480	10,166	17,157	10,374	21,555
Contingency 30%	1,944	3,050	1,545	1,090	1,982
Drilling 15%			1,004	513	1,142
NORAM provided 10%			685	425	921
Subtotal	8,424	13,215	20,392	12,400	25,590
Sales tax 8.4%	708	1110	1,713	1,042	2,150
Subtotal	9,132	14,325	22,105	13,442	27,740
Allied Cost 35%	3,196	5,014	7,737	4,705	9,709
Less NORAM engineering 15%			1,161	706	1,456
Total	*$ 12,328*	*$19,339*	*$28,681*	*$17,441*	*$35,992*
DEWATERING					
Site Work	$20	$20	$20	$20	$20
Structures	780	780	780	780	1,170
Equipment & Mech	2,950	2,950	2,950	2,950	4,425
Electrical/I&C	750	750	750	750	1,125
Testing/Start-up	100	100	100	100	100
Subtotal	4,600	4,600	4,600	4,600	6,840
Contractor Indirects, OH&P 23%	1,053	1,050	1,050	1,050	1,566
Subtotal	5,653	5,650	5,650	5,650	8,406
Contingency 13%	730	730	730	730	1,084
Subtotal	6,383	6,380	6,380	6,380	9,491
Sales tax 8.4%	536	550	550	550	797
Subtotal	6,919	6,930	6,930	6,930	10,288
Allied Cost 35%	2,422	2,430	2,430	2,430	3,601
Total Capital Cost	*$ 9,344*	*$9,360*	*$9,360*	*$9,360*	*$13,889*
Grand Total Capital Expenditure	***$21,668***	***$28,679***	***$38,021***	***$26,781***	***$49,881***

*Thermo-Meso Digestion

3.4.3.2 Operation and Maintenance Cost Estimates

Table 3-5 provides the operation and maintenance cost estimates for the South Treatment Plant alternatives. Excepting Alternative SP-2, the VERTAD™ alternatives have high annual costs for digestion due to the power required for compressed air. Alternative SP-2 has a similar digestion cost as the anaerobic options in part due to the reduced aeration energy and production of significant excess gas resulting from use of excess hot water rather than gas for digester heating. The total annual costs for alternatives SP-1 and SP-2 are lower than those of the anaerobic options due to the factors discussed previously together with lower polymer and biosolids haul and application costs.

Table 3-5: South Treatment Plant Alternatives - O&M Cost Estimate
(All Costs in $1,000's except as indicated)

Category	*TMD*	*Class A*	*Alt SP1*	*Alt SP2*	*Alt SP3*
DIGESTION					
Equipment Maintenance	$83	$101	$102	$91	$106
Operations Labor	296	403	390	376	417
Power					
Fixed	281	314	144	306	243
Variable	123	145	660	188	614
Chemicals			41	41	43
Hot Water Avoided Cost			(27)	(1)	(41)
Gas Sale Net Revenue	(124)	(120)	(84)	(161)	
Total Annual Cost	*$659*	*$843*	*$,1226*	*$841*	*$1,381*
DEWATERING					
Equipment Maintenance	$71	$71	$71	$71	$141
Operations Labor	360	360	360	360	540
Power					
Fixed	2	2	2	2	2
Variable	65	65	58	62	141
Chemicals (polymer)	998	998	678	737	880
Total Annual Cost	*$1,496*	*$1,496*	*$1,168*	*$1,232*	*$1,703*
BIOSOLIDS					
Haul and Application	*$2,147*	*$2,147*	*$1,620*	*$1,762*	*$3,994*
Wet Tons	63,888	63,888	48,205	52,440	78,190
$/WT	33.61	33.61	33.61	33.61	33.61
Dry Tons	15,972	15,972	14,462	15732	35,655
$/DT	134.44	134.44	112.03	112.03	112.03
Total Annual Cost	***$4,302***	***$4,486***	***$4,014***	***$3,835***	***$7,078***

3.4.3.3 Annualized and Present Worth Costs

Table 3-6 provides the present worth cost estimates for the South Treatment Plant alternatives. Alternative SP-2 has a similar present worth as the thermo-meso digestion option which has the lowest cost. The present worth of the avoided costs reflects the year in the future at which point digestion capacity is projected to be exceeded. For the thermo-meso digestion options, the year is 2024; for VERTAD™ options, the years are 2040, 2030, and 2050 for alternatives 1, 2 and 3, respectively.

Table 3-6: South Treatment Plant Alternatives - Present Worth Costs
(All Costs in $1,000's)

Category	*TMD*	*Class A*	*Alt SP1*	*Alt SP2*	*Alt SP3*
Present Worth of Annual Costs	$47,632	$49,740	$44,477	$42,595	$78,200
Present Worth of Capital Costs	21,668	28,679	38,021	26,781	49,881
Subtotal	69,300	78,419	82,498	69,376	128,081
Present Worth of Avoided Costs	(1,146)	(1,146)	(3,043)	(1,961)	(3,855)
Total Present Worth	***$68,154***	***$77,273***	***$79,455***	***$67,415***	***$124,226***

3.4.4 North Treatment Plant

The proposed North Treatment Plant provides the best opportunity for a direct and equivalent comparison between the VERTAD™ process and alternative anaerobic processes. This is the only series of cost estimates in which existing facilities are not used and therefore do not impact the cost effectiveness of the alternatives.

3.4.4.1 Capital Cost Estimates

Table 3-7 provides the capital cost estimates for the North Treatment Plant alternatives. Anaerobic digestion is the lowest capital cost digestion process. The mesophilic anaerobic digestion process is about five percent less costly than the Class A Anaerobic digestion process. The Class A anaerobic digestion process would cost about the same as a VERTAD™ digester (NP-3). The combined VERTAD™ and anaerobic digestion processes (NP-1 and NP-2) have higher capital costs, with NP-2 having the lower cost of the two. Dewatering costs change the ranking of only one alternative. Because the VERTAD™ digester system (NP-3) produces more solids than the other alternatives and must be designed for peaking capacity, the dewatering facility cost is greater. This difference makes the capital cost of digestion and dewatering for NP-3 higher than all alternatives except NP-1.

Table 3-7: North Treatment Plant Alternatives - Capital Cost Estimate (All Costs in $1,000's)

Category	*Anaerobic*		*Alt NP1*	*Alt NP2*	*Alt NP3*
	Base	*Class A*			
DIGESTION					
Site Work	$ 0		$50	$50	$50
Demolition	0				
Anaerobic Digestion	6,500	6,807	3,270	3,806	
Structures	0	0	560	390	700
Equipment & Mech.	0	0	640	540	700
Patent Fee	0	0			
Electrical/I&C	0	0	230	220	220
Testing/Start-up	0	0	incl.	incl.	incl.
NORAM provided equipment and engineering	0	0	3,570	2,890	3,920
Cased Reactors	0	0	3,260	1,620	4,490
Subtotal	6,500	6,807	11,579	9,516	10,080
Contractor Indirects, OH&P 35%	2,275	2,383	1,162	1,349	17
NORAM estimate 11.6%			544	321	709
NORAM provided 5%			179	145	196
Subtotal	8,775	9,190	13,464	11,331	11,002
Contingency 30%	2,633	2,757	2,314	2,437	848
Drilling 10%			379	193	490
NORAM provided 5%			208	172	214
Subtotal	11,408	11,947	16,365	14,132	12,555
Sales tax 8.4%	958	1,004	1,375	1,187	1,055
Subtotal	12,366	12,950	17,739	15,319	13,610
Allied Cost 35%	4,328	4,532	6,209	5,362	4,763
Less NORAM engineering 15%			(931)	(804)	(715)
Total	***$16,694***	***$17,483***	***$23,017***	***$19,877***	***$17,659***
DEWATERING					
Structures	$910	910	910	910	1,365
Equipment & Mech	$1,562	1,562	1,562	1,562	2,135
Testing/Start-up					
Subtotal	$2,472	2,742	2,472	2,472	3,500
Contractor Indirects, OH&P 35%	$865	865	865	865	1,225
Subtotal	$3,338	3338	3,338	3,338	4,725
Contingency 30%	$1,001	1,001	1,001	1,001	1,418
Subtotal	$4,339	4,339	4,339	4,339	6,142
Sales tax 8.4%	$365	365	365	365	516
Subtotal	$4,704	4,704	4,704	4,704	6,658
Allied Cost 35%	$1,646	1,646	1,646	1,646	2,330
Total	***$6,350***	***$6,350***	***$6,350***	***$6,350***	***$8,989***
Grand Total Capital Expenditure	***$23,044***	***$23,833***	***$29,367***	***$26,227***	***$26,648***

In summary, the anaerobic digestion processes are estimated to have lower capital costs than the VERTAD™ alternatives when the costs for digestion and dewatering are combined.

3.4.4.2 Operation and Maintenance Cost Estimates

Table 3-8 provides the operation and maintenance cost estimates for the north treatment plant alternatives. These costs were estimated based on the model for the South Treatment Plant (Ref. 3). Although the actual costs at the future facility will certainly differ, the costs estimated based on the existing facility may be considered relatively conservative.

Table 3-8: North Treatment Plant Alternatives - O&M Cost Estimate (All Costs in $1,000's)

Category	*Anaerobic*		*Alt NP1*	*Alt NP2*	*Alt NP3*
	Base	*Class A*			
DIGESTION					
Equipment Maintenance	$ 67	$114	$87	$75	$53
Operations Labor	188	296	323	323	188
Chemicals			16	16	16
Power	116	205	321	220	263
Hot Water Avoided Cost			(9)	(1)	(13)
Potential Revenue from Gas	(46)	(55)	(33)	(62)	—
Net Energy Cost Subtotal	*70*	*150*	*279*	*157*	*250*
Total Annual Cost	*$326*	*$561*	*$705*	*$572*	*$507*
DEWATERING					
Equipment Maintenance	28	28	28	28	28
Operations Labor	90	90	90	90	180
Power					
Fixed	2	2	2	2	2
Variable	26	23	20	22	46
Chemicals	402	374	235	256	290
Total Annual Cost	*$548*	*$516*	*$374*	*$397*	*$546*
BIOSOLIDS					
Haul and Application	*$769*	*$673*	*$467*	*$509*	*$722*
Wet Tons	27,500	24,000	16,700	18,200	25,800
based on 28 $/WT					
Total Annual Cost	***$1,643***	***$1,749***	***$1,546***	***$1,477***	***$1,775***

The highest electrical energy consumption is estimated for NP-1 because both the anaerobic and VERTAD™ processes have large reactor volumes to mix. NP-3 has the next greatest electrical power consumption. Class A anaerobic and NP-2 have similar electrical power demands. Mesophilic anaerobic digestion has the lowest electrical power requirement. After adjusting for energy value return by methane gas production and excess hot water, the mesophilic anaerobic process has a net energy demand that is less than half of all of the Class A processes. Class A anaerobic and NP-2 have similar net energy demand, although NP-2 is estimated to produce more gas than the anaerobic process. The VERTAD™ process without anaerobic digestion has the highest net energy cost. In summary, the VERTAD™ and anaerobic digestion combinations have the lowest dewatering cost, primarily because of less residual solids and lower polymer demand. The low solids production also results in lower biosolids management costs for these alternatives. In summary, the VERTAD™ and anaerobic digestion combinations (NP-1 and NP-2) have the lowest total operation and maintenance cost.

3.4.4.3 Annualized and present Worth Costs

Table 3-9 provides the present worth cost estimates for the North Treatment Plant alternatives.

Table 3-9: North Treatment Plant Alternatives - Present Worth Costs (All Costs in $1,000's)

Category	*Anaerobic*		*Alt NP1*	*Alt NP2*	*Alt NP3*
	Base	Class A			
Present Worth of Annual Costs	$15,193	$16,827	$14,872	$14,256	$16,440
Present Worth of Capital Costs	23,044	23,833	29,367	26,227	26,648
Total Present Worth	***$38,237***	***$40,660***	***$44,239***	***$40,483***	***$43,088***

Combining the capital costs and operations costs is a present worth analysis shows that the overall low cost alternative is mesophilic anaerobic digestion. For Class A processes, the anaerobic process and the short duration VERTAD™ followed by anaerobic digestion are essentially the same cost.

3.4.5 Cost Sensitivity Analysis

The cost estimates provided above have been developed using design criteria and cost estimating procedures that are judged by the project team to be reasonable based on available information during report preparation in late 2000. These estimates and actual implementation costs would differ if these design and costing assumptions change. The purpose of a cost sensitivity analysis is to document changes in the cost estimate produced by changing individual cost estimating criteria. The cost sensitivity analysis thus provides insight into the cost factors, which have the greatest potential impact on the actual implementation cost. The analysis also provides insight about the potential for reducing costs by investing effort in particular areas. The sensitivity cases were selected to evaluate issues believed to have the greatest potential for impacting the relative desirability of the alternatives.

All of the cost sensitivity comparisons are for the planned new North Treatment Plant. This facility was selected for the sensitivity analysis because the costs are not influenced by existing facilities and their impacts on cost comparison. The following cases have been selected:

Case 1: VERTAD™ *Reactor installation costs are 20 percent higher* - Historically, drilling costs and interest by contractors in competing for contracts have been somewhat variable. It should be assumed that a campaign to interest drilling and mining contractors will be needed to assure low competitive bids. This sensitivity case will provide information on the potential impact of higher installation costs.

Case 2: VERTAD™ *VS reduction is 20 percent less* – Volatile Solids reduction has been somewhat variable over the course of the demonstration testing program. Some of the variability is likely due to mechanical and electrical problems that created unstable operating conditions. However, some of this may be due to normal biological process performance variability. This sensitivity case considers the impact of VS reduction performance that is 20 percent less than expected. Such a change would impact dewatering and biosolid management costs.

Case 3: VERTAD™ *dewatering* - Dewatering effectiveness is critical because of the large impact on haul and application costs. Tests to date have all been bench scale analyses. Although the results have been consistent, application with full-scale equipment could produce different results.

a) *Dewatered cake percent TS reduced by 20 percent.* This case assumes that the product solids content is reduced.
b) *Dewatered cake percent TS increased by 10 percent.* This case assumes that the product solids content is increased.

Case 4: *Energy production and cost* – Energy is a major factor in comparing aerobic and anaerobic systems. Anaerobic processes produce a high value digester gas fuel while consuming energy for process mixing and heating. Aerobic systems use considerable electrical energy for aeration while generating hot water, which is a low grade energy source. Petroleum prices have a major impact on all sources of energy and petroleum derived products such as polymers. Historically, petroleum prices have been variable and stability continues to be uncertain. Therefore, the potential for higher energy prices should be considered when comparing long term costs.

a. *Fuel and polymer price 20 percent and electricity price 100 percent greater* – This case shows how the alternatives cost would react to an increase in energy costs. The value of digester gas was also increased by 20% for this case. Fuel cost for biosolids application was assumed as 15% of the haul cost and 10% of the land application cost, which calculates out as 14% of the total biosolids utilization cost. The cost increase used in this analysis resulted in the following unit costs:
    - Diesel fuel increased from $1.55/gal to $1.86/gal.
    - Digester gas value increased from $0.07/therm to $0.08/therm.
    - Polymer cost increased from $1.80/lb polymer to $2.16/lb polymer.
    - Electricity cost increased from $0.0456/KW-hr to $0.0912/KW-hr.
    - Hot water value increased from $0.02/therm to $0.024/therm.

b. *Anaerobic process gas production 20 percent less* – The only data for gas production in digesters treating VERTAD™ product comes from a bench scale evaluation. This case considers the impact of digester gas production being lower than measured.

Case 5: *Non-conservative VERTAD™ performance assumptions* – Several of the design parameters used to size tankage and equipment and to estimate treatment performance have been selected conservatively based on test results. It is very possible that a full-scale facility design based on further testing and optimized during initial operations would perform better than the demonstration facility. This cost sensitivity case provides insight on the impact of consistent VERTAD™ performance at the higher levels observed during testing to date. The assumed adjustments include:

a) *Oxygen transfer is 65 percent rather than 50 percent* - Transfer rates approaching 61 percent were measured during demonstration testing.
b) *Polymer demand is 15 lb/DT for VERTAD™ product and 20 lb/DT for product that was treated anaerobically after VERTAD™ treatment* – These polymer dosages were found to be appropriate based on bench scale testing.
c) *Dewatered product is 35 percent TS rather than 30 percent* – This solids content was found to be achievable based on bench scale testing.

Case 6: *Class A product used locally* – Producing a Class A product has the potential to open markets within the County service area for utilization of biosolids. These markets are not available to Class B biosolids. GroCo and Tagro biosolids products do participate in the local market on a fairly small scale. The potential exists for substantial reduction in haul and utilization cost. The current average haul and application cost is $28 per wet ton. With sufficient investment in market development it is easy to envision the potential for local haul and minimal tip fee to reduce the average to $12 per wet ton of biosolids. This case considers the impact of this type of market adjustment.

Case 7: *The capital cost of anaerobic digestion is 20 percent greater than estimated* – The capital costs for the North Plant alternatives involving anaerobic digestion are based on the Digester 5 costs for the South Treatment Plant. Since the South Plant already has significant infrastructure in place for this added digester, use of this cost base at the North Plant may underestimate the costs of an anaerobic system. This case considers the impact of anaerobic digester construction costs being 20 percent greater than estimated.

Case 8: *Oxygen demand is 30 percent greater than the typical design value* – Oxygen demand in the VERTAD™ reactor is directly related to the fraction of fats being degraded. Since fat degradation requires 1.6 times the oxygen of carbohydrates and proteins, the oxygen demand for a mix of compounds could be higher than typical design values. The demonstration tests gave some evidence that the biota favors fat and protein over carbohydrates, which could skew the ratio toward higher oxygen demand. This case considers the impact of this factor on alternative costs.

Case 9: VERTAD™ *labor costs reduced by 30 percent* - The demonstration testing program found the VERTAD™ process to be very simple to operate. Without the mechanical and electrical problems experienced with the test facility, the process seems to provide reliable treatment with very little oversight. The cost estimate assumes that VERTAD™ reactor labor is about 75 percent that of an anaerobic digestion system (excluding the SAFT). This case considers the impact of reducing the cost of the VERTAD™ reactor and SAFT labor by 30 percent.

The cost sensitivity analysis was performed only on the costs for the North Treatment Plant. Since this facility would be all new construction, the costs are not influenced by the effects of incorporating existing facilities. The cost impact of the changes can be more directly evaluated. The relative impact of similar changes on alternatives for the other treatment facilities can be approximated from the North Plant comparisons. These comparisons provide useful insights for application of the VERTAD™ technology at the South and West Point Treatment Plants.

The capital cost comparisons for those cases that affect capital costs are provided on Table 3-10. The impact of the change induced for the sensitivity case as expressed by the difference between the base case and the revised cost are shown on Table 3-11.

Table 3-10: Cost Sensitivity Analysis for North Treatment Plant Alternatives - Capital Cost Estimate (All Costs in $1,000's)

Category	*Anaerobic*		*Alt NP1*	*Alt NP2*	*Alt NP3*
	Meso	Class A			
DIGESTION					
VERTAD™ *Reactor*					
Base Case	$0	$0	$ 3,260	$ 1,620	$4,490
1. VERTAD™ Reactor Capital +20%	$0	$0	$ 3,910	$ 1,940	$ 5,390
Anaerobic Digesters					
Base Case	$ 6,500	$ 6,810	$ 3,270	$ 3,810	0
7. Anaerobic Capital +20%	$ 7,800	$ 8,170	$ 3,920	$ 4,570	0
Grand Total Capital Expenditure					
Base Case	$ 23,040	$ 23,830	$29,370	$ 26,230	$26,650
1. VERTAD™ Reactor Capital +20%	$ 23,040	$ 23,830	$30,520	$ 26,800	$28,250
7. Anaerobic Capital +20%	$ 26,380	$ 27,330	$1,080	$ 28,230	$26,650

Table 3-11: Cost Sensitivity Analysis – Capital Cost Impact (All Costs in $1,000's)

Category	*Anaerobic*		*Alt NP1*	*Alt NP2*	*Alt NP3*
	Meso	Class A			
1. VERTAD™ Reactor Capital +20%	$ -	$ -	$1,150	$ 570	$ 1,600
7. Anaerobic Capital +20%	$ 3,340	$ 3,500	$1,710	$ 2,000	$ -

As indicated these capital costs can have a significant effect on the alternative comparison. In particular, the anaerobic cost increase would make alternative NP3 more attractive from a capital cost basis.

The operation and maintenance cost comparisons for those cases that effect O&M costs are provided on Table 3-12. The impact of the O&M cost changes are shown on Table 3-13.

Table 3-12: Cost Sensitivity Analysis for North Treatment Plant Alternatives - O&M Cost Estimate (All Costs in $1,000's)

Category	**Anaerobic**		**Alt NP1**	**Alt NP2**	**Alt NP3**
	Meso	Class A			
DIGESTION					
Power - Variable					
Base Case	$ 24	$ 28	$ 227	$ 126	$ 202
2. VERTAD™ VSR -20%	$ 24	$ 28	$ 187	$ 106	$ 162
4a. Energy Cost +100%	$ 49	$ 57	$ 454	$ 252	$ 404
5. VERTAD™ Optimum performance	$ 24	$ 28	$ 180	$ 103	$ 155
8. O_2 Demand +30%	$ 24	$ 28	$ 313	$ 169	$ 288
Gas Sale Net Revenue					
Base Case	$ (46)	$ (55)	$ (33)	$ (62)	
2. VERTAD™ VSR -20%	$ (46)	$ (55)	$ (40)	$ (67)	
4a. Energy Cost +20%	$ (56)	$ (66)	$ (40)	$ (74)	
4b. VERTAD™ Gas Production -20%	$ (46)	$ (55)	$ (27)	$ (49)	
5. VERTAD™ Optimum performance	$ (46)	$ (55)	$ (33)	$ (62)	
Labor					
Base Case	$ 188	$ 296	$ 323	$ 323	$188
9. VERTAD™ labor -30%	$ 188	$ 296	$ 282	$ 282	$148
DEWATERING					
Power - variable					
Base Case	$ 26	$ 23	$ 20	$ 22	$ 46
4a. Energy Cost +100%	$ 53	$ 45	$ 39	$ 43	$ 93
Polymer					
Base Case	$ 402	$ 374	$ 235	$ 256	$ 290
2. VERTAD™ VSR -20%	$ 402	$ 374	$ 252	$ 261	$ 318
4a. Energy Cost +20%	$ 482	$ 448	$ 282	$ 307	$ 348
5. VERTAD™ Optimum performance	$ 402	$ 374	$ 188	$ 204	$ 218
BIOSOLIDS					
Base Case	$ 769	$ 673	$ 467	$ 509	$ 722
2. VERTAD™ VSR -20%	$ 769	$ 673	$ 501	$ 520	$ 790
3a. %TS -20%	$ 769	$ 673	$ 584	$ 636	$ 903
3b. %TS +10%	$ 769	$ 673	$ 425	$ 463	$ 657
4a. Energy Cost +20%	$ 791	$ 692	$ 480	$ 523	$ 742
5. VERTAD™ Optimum performance	$ 769	$ 673	$ 400	$ 436	$ 619
6. Class A Biosolids -50%	$ 769	$ 337	$ 234	$ 254	$ 361
Total Annual Cost					
Base Case	$ 1,642	$ 1,749	$ 1,546	$ 1,477	$ 1,775
2. VERTAD™ VSR -20%	$ 1,642	$ 1,749	$ 1,552	$ 1,469	$ 1,835
3a. %TS -20%	$ 1,642	$ 1,749	$ 1,663	$ 1,604	$ 1,956
3b. %TS +10%	$ 1,642	$ 1,749	$ 1,504	$ 1,431	$ 1,710
4a. Energy Cost +20% & 100%	$ 1880	$ 2061	$ 1940	$ 1773	$ 2162
4b. VERTAD™ Gas Production -20%	$ 1,642	$ 1,749	$ 1,554	$ 1,490	$ 1,775
5. VERTAD™ Optimum performance	$ 1,642	$ 1,749	$ 1,386	$ 1,330	$ 1,553
6. Class A Biosolids -50%	$ 1,642	$ 1,413	$ 1,313	$ 1,223	$ 1,414
8. O_2 Demand +30%	$ 1,642	$ 1,749	$ 1,633	$ 1,520	$ 1,862
9. VERTAD™ labor -30%	$ 1,642	$ 1,749	$ 1,506	$ 1,437	$ 1,735

Table 3-13: Cost Sensitivity Analysis – Total O&M Cost Impact (All Costs in $1,000's)

Category	Anaerobic		Alt NP1	Alt NP2	Alt NP3
	Meso	Class A			
2. VERTAD™ VSR –20%			$ 6	$ (8)	$ 60
3a. %TS –20%			$ 117	$ 127	$ 181
3b. %TS +10%			$ (42)	$ (46)	$ (65)
4a. Energy Cost +20% & 100%	$ 238	$ 312	$ 394	$ 296	$ 387
4b. VERTAD™ Gas Production –20%			$ 8	$ 13	$ -
5. VERTAD™ Optimum performance			$ (160)	$ (147)	$ (222)
6. Class A Biosolids –50%		$ (336)	$ (233)	$ (254)	$ (361)
8. O_2 Demand +30%			$ 87	$ 43	$ 87
9. VERTAD™ labor -30%			$ (40)	$ (40)	$ (40)

The operation and maintenance factors that have the greatest impact on costs are dewatering, energy, VERTAD™ performance and Class A product markets.

The present worth cost comparisons are provided on Table 3-14. The impact of the present worth cost changes are shown on Table 3-15.

Table 3-14: Cost Sensitivity Analysis for North Treatment Plant Alternatives – Total Present Worth Costs (All Costs in $1,000's)

Category	Anaerobic		Alt NP1	Alt NP2	Alt NP3
	Meso	Class A			
1. VERTAD™ Reactor Capital +20%	38.2	40.7	45.4	41.1	44.7
2. VERTAD™ VSR –20%	38.2	40.7	44.9	40.6	43.6
3a. %TS –20%	38.2	40.7	45.2	41.6	44.6
3b. %TS +10%	38.2	40.7	43.9	40.1	42.6
4a. Energy Cost +20% & 100%	40.5	43.8	47.8	43.3	46.5
4b. VERTAD™ Gas Production – 20%	38.2	40.7	44.3	40.6	43.1
5. VERTAD™ Optimum performance	38.2	40.7	42.9	39.3	41.3
6. Class A Biosolids –50%	38.2	37.8	42.3	38.4	40.1
7. Anaerobic Capital +20%	41.6	44.2	46	42.5	43.1
8. O_2 Demand +30%	38.2	40.7	45	40.9	43.8
9. VERTAD™ labor -30%	38.2	40.7	43.8	40	42.7
Base Case	38.2	40.7	44.2	40.5	43.1

Table 3-15: Cost Sensitivity Analysis – Total Present Worth Cost Impact (All Costs in $1,000,000's)

Category	***Anaerobic***		***Alt NP1***	***Alt NP2***	***Alt NP3***
	Meso	Class A			
1. VERTAD™ Reactor Capital +20%	0	0	1.2	0.6	1.6
2. VERTAD™ VSR –20%	0	0	0.7	0.1	0.5
3a. %TS –20%	0	0	1	1.1	1.5
3b. %TS +10%	0	0	-0.3	-0.4	-0.5
4a. Energy Cost +20% & 100%	2.3	3.1	3.6	2.8	3.4
4b. VERTAD™ Gas Production – 20%	0	0	0.1	0.1	0
5. VERTAD™ Optimum performance	0	0	-1.3	-1.2	-1.8
6. Class A Biosolids –50%	0	-2.9	-1.9	-2.1	-3
7. Anaerobic Capital +20%	3.4	3.5	1.8	2	0
8. O_2 Demand +30%	0	0	0.8	0.4	0.7
9. VERTAD™ labor -30%			-0.4	-0.5	-0.4

Table 3-16 provides an overall comparison of the impact of the sensitivity cost changes on capital, operations and present worth costs.

Table 3-16: Cost Sensitivity Analysis – Cost Impact (% Change from Base Case Cost)

Category	Anaerobic		Alt NP1	Alt NP2	Alt NP3
	Meso	Class A			
Capital Cost					
1. VERTAD™ Reactor Capital +20%	0%	0%	4%	2%	6%
7. Anaerobic Capital +20%	14%	15%	6%	8%	0%
O&M Cost					
2. VERTAD™ VSR –20%			0%	-1%	3%
3a. %TS –20%			8%	9%	10%
3b. %TS +10%			-3%	-3%	-4%
4a. Energy Cost +20% & 100%	14%	18%	25%	20%	22%
4b. VERTAD™ Gas Production – 20%			1%	1%	0%
5. VERTAD™ Optimum performance			-10%	-10%	-13%
6. Class A Biosolids –50%		-19%	-15%	-17%	-20%
8. O_2 Demand +30%			6%	3%	5%
9. VERTAD™ labor -30%			-3%	-3%	-2%
Present Worth Cost					
1. VERTAD™ Reactor Capital +20%	0%	0%	2.7%	1.5%	3.7%
2. VERTAD™ VSR –20%	0%	0%	2%	0%	1%
3a. %TS –20%	0%	0%	2%	3%	3%
3b. %TS +10%	0%	0%	-1%	-1%	-1%
4a. Energy Cost +20% & 100%	6.0%	7.6%	8.1%	6.9%	7.9%
4b. VERTAD™ Gas Production – 20%	0%	0%	0%	0%	0%
5. VERTAD™ Optimum performance	0%	0%	-3%	-3%	-4%
6. Class A Biosolids –50%	0%	-7.1%	-4.3%	-5.2%	-6.8%
7. Anaerobic Capital +20%	8.9%	8.6%	4.1%	4.9%	0%
8. O_2 Demand +30%	0%	0%	2%	1%	2%
9. VERTAD™ labor -30%			-1%	-1%	-1%

As a method of prioritizing the significance of the sensitivity cases, a threshold of five percent change is considered moderate and a change of 10 percent is considered significant. Using these criteria the primary factors identified by this analysis are:

Primary Cost Variables (greater than 10 percent change)

- Anaerobic digestion-capital cost
- Dewatering percent TS-operating cost
- VERTAD™ optimum performance-operating cost
- Class A biosolids operating cost
- Increased energy costs

Secondary Cost Variables (greater than 5 percent change)

- VERTAD™ oxygen demand

Only the Class A biosolids utilization fee would result in greater than five percent change in the present worth cost for any alternative.

3.5 Comparison of Alternative Facility Space Requirements

The subsurface configuration used by the VERTAD™ requires less space than traditional anaerobic digestion processes. Figure 3-10 provides a comparison of space requirements for the North Treatment Plant alternatives. For the layouts shown, the space needed for alternative NP3 is less than a third of the requirement for the mesophilic anaerobic digestion alternative. Similar space reductions could be expected at the West Point and South Treatment Plants, but would require removal of existing digesters.

3.6 Implementation Strategy

The cost comparison in this evaluation indicated that the VERTAD™ technology should be considered for full-scale implementation at King County facilities. An implementation strategy for the VERTAD™ technology should include additional testing using the existing demonstration facility. This would allow refinement of design criteria in preparation for a full-scale installation. The combined results of the performance testing and cost estimates that have been developed during this evaluation indicate that any further evaluation of the technology should focus on three primary areas:

1. Dewatering of the VERTAD™ product to determine reliable expected performance for percent Total Solids and polymer demand. Considering the potential benefits of sulfuric acid flotation (SAFT), dewatering should be evaluated for the following configurations:
   a) VERTAD™ followed by SAFT
   b) VERTAD™ followed by SAFT followed by anaerobic digestion (mesophilic and thermophilic)
   c) Anaerobic digestion (mesophilic and thermophilic) followed by VERTAD™ followed by SAFT
2. Linked anaerobic and VERTAD™ process configurations appear to be the most cost effective based on the limited operational evaluation conducted to date. These configurations should be evaluated in more depth.
3. Class A product market potential should be evaluated to determine how much Class A product could be used locally. Development of a local market for Class A product would dramatically reduce hauling costs. Competitive pressures created by these markets may also result in reduced costs for the Class B product markets. Development of this market may also impact the consideration of the Centridry process. Any differences in market potential for Class A products derived from VERTAD™ and anaerobic processes should also be evaluated. A well developed Class A product market may also impact the cost effectiveness of the Centridry process. Centridry is most attractive for long distance hauling because of the weight and volume reduction. The availability of local markets for Class A product would significantly reduce the potential benefit associated with Centridry.

Provided that further testing supports the cost effectiveness of the process, it is recommended that the County consider construction of a full-scale VERTAD™ facility at the South and North Treatment Plants. Depending on the results of further testing to determine the preferred sequence, the VERTAD™ process at the South Treatment Plant could be designed to either:

- Process all solids fed to one of the four existing digesters.
- Process sufficient solids from an anaerobic digester to feed one centrifuge.
- Process through short duration VERTAD™ feeding all four digesters.

In either instance the VERTAD™ process could be designed to provide sufficient excess heat to maintain operating temperatures in the anaerobic digesters without the use of gas.

Demonstration and Evaluation of

VERTAD™
Aerobic Thermophilic Digestion Process

Section 4

VERTAD™ Development Assessment (Summary of Development Status)

The VERTAD™ Demonstration Project has been very productive in defining the performance capabilities and design and operating criteria for the process. This section is provided as a summary of the current state of knowledge about the VERTAD™ process.

Each topic is discussed relative to the significance of the topic, current knowledge, and additional information that would be desirable prior to technology implementation. Also provided is an assessment of additional information that would be desirable to have prior to implementing the VERTAD™ process on a large-scale basis. As always the amount of knowledge needed prior to full-scale implementation is a matter of judgement, and is subject to adjustment. It is also possible to design a full-scale facility with sufficient flexibility to allow successful design and operation with less than complete knowledge.

4.1 Process Simplicity and Stability

Process simplicity and stability are highly desirable characteristics of any wastewater treatment process. Simplicity refers to the number of control points, the frequency that they must be adjusted and the complexity of tests and procedures needed to determine when and to what extent adjustment is needed. Stability refers to response of the process to normal variations in feed solids and flow and other factors that might affect operations. Stability also refers to the potential for upset and the ability of the process to recover from upset inducing conditions.

4.1.1 Current Knowledge

The VERTAD™ process is quite simple both mechanically and electrically. The demonstration unit is actually more complex than a continuously fed full-scale system because of the need for batch feeding. A full-scale system may have a control system to vary aeration rate with demand. The VERTAD™ process uses only very basic and standard equipment that is normally associated with wastewater solids digestion. The greatest operations challenge associated with a full-scale VERTAD™ may be removal of heat to maintain the desired operating temperatures throughout the reactor.

Use of the sulfuric acid flotation system would add complexity to the VERTAD™ process. Combining VERTAD™ and anaerobic digestion will result in a digestion process that is more complex than either component. However, there appear to be sufficient synergistic benefits from combining the processes that the combined process will not be excessively complex. These benefits are primarily associated with heat recovery and transfer.

The VERTAD™ process has demonstrated repeatedly that it is very stable and resistant to upset. The demonstration facility experienced a range of mechanical and electrical problems that repeatedly altered feed patterns and stressed the biota in a variety of ways. In all cases the biological system demonstrated stability and rapid recovery.

4.1.2 Knowledge Needed Before Implementation

The stability and simplicity of the VERTAD™ process has been well demonstrated. The simplicity of VERTAD™ operated together with the sulfuric acid flotation thickening (SAFT) system and/or anaerobic digestion has not been demonstrated at an operational scale.

4.2 Anaerobic Linking

Combining VERTAD™ and anaerobic digestion was found to be the most cost-effective approach for implementing the VERTAD™ technology. This combination had the lowest operation and maintenance cost of the Class A alternatives considered, including anaerobic processes.

4.2.1 Current Knowledge

Current understanding of the combined process comes from operation of bench scale anaerobic digesters treating product derived from operating VERTAD™ at a four-day residence time. The results of these tests were promising with regard to overall volatile solids reduction, anaerobic gas production and scum and foam control.

4.2.2 Knowledge Needed Before Implementation

Continuous operation of a pilot-scale, combined VERTAD™ / anaerobic digestion system would be very desirable. Optimization of the respective retention times to achieve sufficient solids destruction, maximize gas production, and minimize the energy requirement could be conducted at the pilot-scale. In addition to the combination tested at bench scale, the following configurations would be desirable to test:

1. *Short duration VERTAD™ followed by anaerobic digestion.* This combination was found to be the most cost effective VERTAD™ process in the alternative analysis. Optimizing detention times in the respective reactors should be a primary objective.
2. *Short duration anaerobic digestion followed by VERTAD™.* This combination may prove to be more cost effective due to increased gas production and better sequencing of constituent degradation. Again, the optimum detention times in the respective reactors are unknown.

Continuous operation of the SAFT process in conjunction with anaerobic digestion is a significant economic factor. Thickening between processes would significantly reduce the required capacity of the anaerobic reactors. Pilot-scale testing could indicate the appropriate limit on total solids content of the thickened feed to an anaerobic digester. This is a particularly important issue with the short duration VERTAD™ reactor because sidestream quality may have a major impact on the liquid stream processes.

4.3 Residence Time and Feed / Reactor TS Content

Residence time in the reactor directly impacts the size and cost of the VERTAD™ reactor. Similarly, since the reactor is sized based on organic loading, the feed solids content has a direct relationship to the size and cost of the reactor.

4.3.1 Current Knowledge

The demonstration tests indicate that oxygen transfer can be limited by excessively high solids content in the reactor. Oxygen transfer and feed solids content must be considered in sizing the VERTAD™ reactor. Tests to date indicate that a four-day detention time is needed to reliably provide VAR compliance of 38 percent VS reduction. However, combining VERTAD™ with anaerobic digestion has been shown with bench scale tests to readily comply with the VAR requirement.

4.3.2 Knowledge Needed Before Implementation

The relationship between reactor solid content (or some other measure of viscosity) and oxygen transfer needs to be better defined in order to optimize the design of the VERTAD™ reactor.

See the above discussion of issues related to combining VERTAD™ and anaerobic digestion.

4.4 Class A Pathogen Control

The economic analysis indicates significant cost savings potential associated with local use of a Class A biosolids product. VERTAD™ operates at temperature consistent with production of a Class A product.

4.4.1 Current Knowledge

The VERTAD™ process relies on the heat generated biologically to maintain sufficient temperature for Class A pathogen control. The relative amounts of heat generated and heat lost to the surroundings are dependent on the reactor diameter and the resultant surface area to volume ration. As the diameter of the reactor is increased, the volume increases by a factor of the diameter squared while the surface area increases linearly. Thus, the reactor volume (which determines the amount of organic material destroyed and the heat generated) increases at a greater rate than the surface area (which determines heat loss). The demonstration facility needs supplemental heat to maintain Class A required operating temperatures. However, calculations based on estimated heat loss and generation for this facility show that the system would be "autothermal" (no supplemental heat required) at a diameter of between three and four feet when operating at 55°C. Insulating grout could also be used to achieve autothermal conditions in smaller reactors. Therefore, full-scale reactors for King County facilities (10 to 13 foot diameter) would generate a significant amount of excess heat which could need to be removed to prevent overheating of the reactor. It is well established on this basis that the VERTAD™ process will easily generate temperatures sufficient to comply with Class A requirements.

Equally critical to temperature for Class A compliance is the time of contact. Assured time of contact requires special precautions to prevent short circuiting and resulting pathogen pass through. Mixing tests in the VERTAD™ reactor indicate that mixing in the upper two zones occurs in a period of several hours. This zone alone is insufficient to ensure that short-circuiting is prevented in the reactor (except possibly for operation at very high temperatures of 70 to 75°C). VERTAD™, therefore, relies on detention time in the unmixed lowest zone of the reactor, the soak zone, to provide required contact time. Tracer measurements have shown that this zone operates in plug flow and that a properly designed VERTAD™ with a soak zone will meet the contact time requirements for Class A compliance.

There has been no evaluation of the potential for use of Class A VERTAD™ product in local organic product markets. An evaluation of organic waste management recently completed for the County identified the topsoil market as a large potential market for organic waste derived products. This market is also being evaluated by the County for Centridry compost product. VERTAD™ and combined VERTAD™ / anaerobic products would be expected to have potential for use in this and other markets. In addition, the VERTAD™ product might be processed by GroCo, Inc. at a lower fee than currently because of the higher solids content and Class A designation.

4.4.2 Knowledge Needed Before Implementation

Marketing information to be collected by the County regarding using Centridry product as a constituent of manufactured topsoil will be useful for determining if that is a potential market for VERTAD™ Class A product. Specific comparison of VERTAD™ products with the organic matter characteristics desired by organic product markets with potential interest in Class A biosolids would also be desirable. In order to sell the product into most markets it will likely be necessary to perform a range of tests and demonstrations including growth trials.

4.5 Product Suitability for Beneficial Use

The characteristics of the VERTAD™ product are important for incorporation into the current land application program or in new Class A markets. Appearance, odor, nutrient content, and application characteristics are all of importance.

4.5.1 Current Knowledge

Very limited information is available about the character of VERTAD™ product after dewatering. The visual appearance is lighter in color than the currently produced anaerobic cake with the appearance of a higher fiber content.

4.5.2 Knowledge Needed Before Implementation

Information about the character of the VERTAD™ product relative to market quality and aesthetics should be developed along with any dewatering testing.

4.6 Thickening/Dewatering/Centridry

Thickening and dewatering are critical to the economic viability of the VERTAD™ process. Taking VERTAD™ product to a drier state with Centridry could further improve the cost savings available with VERTAD™.

4.6.1 Current Knowledge

Thickening with the SAFT process has been demonstrated on a batch, bench scale basis. The concepts and equipment used for this process are common in the wastewater industry. The difference is floating with pH shift induced CO_2 bubbles rather than dissolved air as used in the DAFT process.

Dewatering of VERTAD™ product has been conducted in the field and at the laboratories of dewatering equipment manufacturers. In the Andritz laboratory, the thickened VERTAD™ product dewatered to between 31-34 percent with a 99.5 percent capture efficiency on a polymer dose of 14 lbs/ton. VERTAD™ product that was not thickened produced similar cake solids content but required about three times the polymer dose. The only full-scale dewatering of VERTAD™ product was a one-day test with the Centridry process. Although only indirect observation of dewatering is possible with this equipment, the conclusion of the test was that VERTAD™ product dewatered very well but had a relatively high polymer demand

It is recommended that dewatering with the Fournier rotary press and FKC press be evaluated for VERTAD™ product. These dewatering devices have been demonstrated to be very effective at dewatering high fiber solids, as the VERTAD™ product appears to be. The product solids content from this equipment may be significantly greater than test results to date.

4.6.2 Knowledge Needed Before Implementation

More data, particularly from continuous dewatering operations, would be very desirable to confirm the potential for high solids content product, high capture efficiency, and low polymer requirements. More information on dewatering of the product from combining VERTAD™ and anaerobic processes is also needed.

Pre-dewatering thickening in a SAFT system may also improve the odor and utility of the dewatered product. Tests to document any changes would be useful in determining the market value of the Class A product.

4.7 Microbial Degradation of Organic Constituents

Fats, oils, and grease (FOG) have significantly different chemical characteristics compared to protein and carbohydrates with regard to digester performance. FOG requires about 2.5 times more oxygen and releases the same additional energy per pound of solids degraded. This becomes significant if fat is preferentially degraded relative to carbohydrates and proteins at any operating conditions.

4.7.1 Current Knowledge

Monitoring results indicate that the VERTAD™ process selectively degrades FOG at higher rates than carbohydrates and protein (80 percent for FOG vs. 50 percent for organic nitrogen and 40 percent for VS). Protein appears to degrade at higher rates than carbohydrates. By difference, considering the overall VS reduction, carbohydrates would be assumed to degrade at rates below the VS reduction. This is partially substantiated by the fibrous appearance of the dewatered product.

In addition, supplemental feed testing shows oil consumption is very rapid and only limited by oxygen transfer. Supplemental feed of sugar, however, showed a much slower consumption and oxygen demand below the transfer limit.

4.7.2 Knowledge Needed Before Implementation

Since preferential degradation of fat can impact design criteria for aeration and heat removal, determination of appropriate design values is important for full-scale implementation. The issues raised by previous test results indicates the need for more in depth analysis to determine the relative consumption of these primary organic matter types.

This may be especially important for the combined VERTAD™ and anaerobic systems. Preferential fat degradation in the short duration VERTAD™ reactor could require greater oxygen transfer and provide greater VS reduction than anticipated. Removal of FOG from the anaerobic process may also provide operational benefits for that process.

4.8 Energy Release, Loss, and Consumption

The release of energy during degradation, loss of that energy to the subsurface geology surrounding the reactor, and the consumption of energy by aeration compressors are important design and economic factors. Understanding these issues is important for the design of a facility that performs as expected.

4.8.1 Current Knowledge

Energy loss to adjacent geology and surface air has been measured at the demonstration site on several occasions when biological heat release would not be significant. These measurements have shown significant variability. Much of the variability is thought to be derived from changes in groundwater flow around the reactor. However, these variations do not match expected seasonal changes in flow or induced flow that may have been a factor during nearby construction. Heat loss measurement is complicated by vertical changes in the subsurface profile from gravel at the surface to saturated sand and gravel to rock.

Energy release from degradation was discussed in the previous item. Measurement of degradation energy release has been by difference using the estimated heat loss to the environment, boiler heat addition, feed energy demand, and reactor temperature. Since these calculations depend on the variable loss data, the accuracy of these calculations is also subject to variability. Degradation heat release has also been estimated using degradation rates for FOG and organic nitrogen (protein). COD reduction data is also a reasonable method of estimating biological energy release in an aerobic system. These methods have not provided consistent results to date. Fermentation of carbohydrates may be occurring in the VERTAD™ reactor based on low measured ORP. Fermentation rather than oxidation may change the energy release characteristics and explain some of the variability. An effective method of measuring oxygen demand and heat release has been through supplemental addition of known constituent organic feed. Both vegetable oil and high fructose syrup have been added. The impact of these additions were directly and accurately measured. Oil addition demonstrated rapid oxygen consumption and energy release consistent with theory. The addition of sugar produced results that were less clear.

The electrical energy requirement for aeration is a primary energy and cost factor. Much of the energy input by the compressor can be recovered as hot air or hot water. This issue is discussed in more detail in the following item.

4.8.2 Knowledge Needed Before Implementation

Continued measurement of heat loss from the reactor and developing better understanding of degradation energy release would provide desirable data to be used for full-scale design. The heat loss to groundwater needs to be understood or an insulated reactor used at sites with significant groundwater impact.

Expanded use of the supplemental feed test is recommended to better understand degradation rates, energy release and oxygen demand for a range of fats, proteins, and carbohydrates.

4.9 Aeration / Oxygen Transfer

The ability of VERTAD™ to transfer oxygen for use by the thermophilic organisms is critical to cost effective operation.

4.9.1 Current Knowledge

Oxygen transfer has been directly and accurately measured for the VERTAD™ reactor. The process is ideal for measurement of oxygen transfer. This characteristic provides a very accurate and powerful process control tool.

Measurements in the demonstration reactor indicate that higher solids levels impede oxygen transfer and limit the VS reduction performance of the process. This is theorized to result from viscosity effects on the transfer of oxygen from gas into the liquid. This solids content limitation is an economic factor because it increases the size of reactor required when feed solids are thick.

Oxygen transfer rates as high as 61 percent have been measured in the reactor following the addition of supplemental oil. The highest transfer rates measured during feeding of thickened solids was 53 percent.

4.9.2 Knowledge Needed Before Implementation

It would be desirable to better define the relationship between reactor solids content and oxygen transfer capability.

Application of the demonstration reactor data to selecting an air release configuration for a full-scale facility also warrants additional evaluation. The efficient transfer of oxygen may be affected by the placement of aeration diffusers in the reactor (single versus multiple injection array). Experience that NORAM has developed for large, full-scale VERTREAT™ (activated sludge version of vertical reactor) applications provide additional information on this issue.

4.10 Vector Attraction Reduction

The VERTAD™ process complies with the VAR requirement by reducing the volatile solids content by 38 percent or greater. In addition to this basic regulatory requirement, VS reduction also effects the quantity of solids that must be dewatered, hauled, and utilized. VS reduction is therefore an important economic factor.

4.10.1 Current Knowledge

VERTAD™ performance when operating at a four-day detention time has exceeded 42 percent provided that the feed was diluted or the operating temperature greater than 65°C. This relatively low VS destruction compared to anaerobic systems appear to be primarily the result of poor degradation of paper fiber. Degradation of FOG and organic nitrogen has been routinely significantly greater than 44 percent. Since it is also believed that vector attracting carbohydrates are also readily degraded, there is strong evidence that the VERTAD™ process readily meets the intent of the regulation (reducing attraction to vectors) while marginally meeting the numerical standard.

Combining VERTAD™ with anaerobic digestion eliminates this issue because the combined process greatly exceeds the VS reduction requirement.

4.10.2 Knowledge Needed Before Implementation

As part of future testing, the use of the oxygen uptake test as an alternative to VS reduction as the accepted measure of VAR for VERTAD™ should be pursued.

4.11 Mass Balance

A process mass balance is a basic tool for process operation and control. High rate, high temperature, aerated systems present a special challenge because of the sensitivity of evaporation and moisture removal to the operating temperature.

4.11.1 Current Knowledge

Current instrumentation on the reactor has allowed development of an accurate mass balance for the process.

4.11.2 Knowledge Needed Before Implementation

Continued mass balance tracking in any future testing is desirable.

4.12 Sidestream Characteristics

Liquid sidestreams from solid – liquid separation processes have potentially significant effects upon return to the wastewater treatment facilities liquid treatment processes. Solids, BOD, and nutrients are parameters of interest.

4.12.1 Current Knowledge

Only limited data on solid content and nutrients are available from small scale batch testing.

4.12.2 Knowledge Needed Before Implementation

Sidestream quality and quantity monitoring should be included as part of any future thickening and dewatering testing.

4.13 Operating Temperature

The temperature at which the VERTAD™ process is operated has significant impacts on degradation rates, Class A compliance, heat loss, water loss, and pH balance. In short, the operating temperature is a critical operating criteria.

4.13.1 Current Knowledge

The demonstration project has shown that the operating temperature in the reactor can be controlled with minimal variation through heat addition. It is feasible to accomplish a similar level of temperature stability in a full-scale facility via heat removal. However, due to heat loss variations in the demonstration reactor and lack of clarity about the heat release from degradation, it is currently not clear which portion of the reactor will require the highest level of cooling. Reactor mixing may adequately equalize temperature gradients. If not, this uncertainty can be overcome through the use of flexible heat removal systems that allow heat removal from specific areas of the reactor as required.

Water loss and deposition at various points in the VERTAD™ process are well understood and predictable.

4.13.2 Knowledge Needed Before Implementation

Additional information on heat generation in each zone of the reactor would be useful for optimizing the design of heat removal and recovery systems. Heat generation and mixing in the slowly mixed zone are of particular interest.

4.14 Reactor Mixing

Mixing is important relative to contact time for Class A compliance and for optimizing the potential for VS reduction.

4.14.1 Current Knowledge

Mixing in the demonstration reactor is well understood based on tracer studies. As expected, the upper zone is highly mixed. The "plug flow" zone was found to be slowly mixed in a matter of four to eight hours depending on the aeration rate. Translation of this information to a full-scale reactor with nine to 13 foot diameter is not clear. Some relevant research and theory indicates that mixing in a larger reactor should be similar. However, this has not been demonstrated on a full-scale basis.

The tracer studies indicated minimal mixing in the lower "soak" zone at the bottom of the reactor.

4.14.2 Knowledge Needed Before Implementation

None, although it would be prudent to design the initial full-scale reactor to allow easy modification of the aeration system to control mixing in the lower zone.

It would also be desirable to evaluate mixing in the lower "soak" zone in more detail to provide assurance that density currents or some other mechanism would not result in excessive mixing for Class A compliance.

4.15 Odor / Off Gases

Odor and off gas control is an issue with any solids handling facility.

4.15.1 Current Knowledge

Tests and observations during testing indicate that the off gas from the VERTAD™ process is odorous but treatable with biofiltration. Further, it was demonstrated that the bulk of odor was generated by bubbling the offgas through the undegraded feed solids rather than from the VERTAD™ itself. Alternate feed and energy recovery strategies may be used to mitigate this odor.

4.15.2 Knowledge Needed Before Implementation

Additional performance data should be collected as part of future testing. It would also be desirable to test the performance of the water based biofiltration system proposed for full-scale application.

4.16 Potential for Use in Food Waste Management

The County is considering alternatives for managing food waste and other organics generated in the service area. VERTAD™ is a process that has potential for converting these source-separated or sewer-transported food waste materials into a recyclable product.

4.16.1 Current Knowledge

Economic analyses indicate that it is cost effective to source separate and process food waste from large generators. It also appears to be cost effective (although subject to continuing debate) to divert food waste from the solid waste stream to the sewer system. The primary problem in both cases is a lack of processing capacity. Treatment plants are not sized for a massive influx of food waste that is currently being landfilled and most local composters are not permitted to accept post-consumer food waste, meat or oily wastes. Experience operating the VERTAD™ process indicates that it would readily accept and process all food waste except associated paper products. VERTAD™ could be used as added capacity to treat the additional load from sewer transported food waste or source separated food waste transported directly to the VERTAD™ process.

4.16.2 Knowledge Needed Before Implementation

Macerated and liquified source separated food waste could be fed to the reactor to document performance treating this material.

4.17 Safety

Safety concerns are a paramount concern. The only unusual safety characteristic of the VERTAD™ process are the high operating temperature and the use of moderately high pressure air.

4.17.1 Current Knowledge

The VERTAD™ operating temperatures are common in ATAD and thermophilic anaerobic systems and composting. These operating temperatures, while requiring established safe operating procedures, are not considered an impediment to successful operation of the process. Hot liquids and gases were not a problem during operation of the demonstration facility.

4.17.2 Knowledge Needed Before Implementation

No additional information is needed.

4.18 Corrosion

4.18.1 Current Knowledge

Corrosion has not been found to be a problem with the demonstration reactor or other similar deep biological reactors.

4.18.2 Knowledge Needed Before Implementation

None, although corrosion rate measurement should be formalized if this is a continuing concern.

4.19 Earthquake Damage

There is an obvious concern with earthquake damage to subsurface reactors.

4.19.1 Current Knowledge

Evaluation of the potential for earthquake damage was completed before the demonstration reactor was constructed. The evaluation determined that the deep reactors were less likely to experience earthquake damage than surface tankage.

4.19.2 Knowledge Needed Before Implementation

None

4.20 Groundwater / Subsurface effects

Operation of a VERTAD™ reactor will result in heating of the subsurface geology and associated groundwater.

4.20.1 Current Knowledge

There are no identified issues associated with heating of the subsurface geology.

4.20.2 Knowledge Needed Before Implementation

If heating of the subsurface or groundwater becomes a concern, then future construction could include insulation to minimize impact. The extent to which groundwater is heated and cools as it moves past the reactor could also be monitored by installation of monitoring wells.

4.21 Deposition/Struvite

Crystaline deposits can create operation and maintenance concerns.

4.21.1 Current Knowledge

Crystaline deposits were observed on the walls of the VERTAD™ reactor head tank on occasion. The deposits did not affect operations or require maintenance.

4.21.2 Knowledge Needed Before Implementation

Continue to observe and document deposits and identify the compound if it interferes with operations.

4.22 Reactor Construction Bidding Competitiveness

VERTAD™ reactor construction is a large portion of the construction cost. Cost control is needed to obtain a reasonable price for construction.

4.22.1 Current Knowledge

Experience indicates that obtaining competitive bids for reactor drilling is not always assured. This is an even greater concern with the large nine to 13-foot diameter reactors. Few drilling companies are capable of drilling holes of this size and none are located in this part of the country. For larger diameter reactors, mining techniques can also be used. This alternative provides improved competitiveness and assurance of a reasonable bid.

4.22.2 Knowledge Needed Before Implementation

Development and maintenance of contact with contractors capable of completing reactor installation is recommended.

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APPENDIX A

University of Washington Study

Effect of VERTAD™ Pre-Treatment on Anaerobic Digestion

By Jenny Yoo

Introduction and Objectives

The Technology Development Program, formerly known as the Applied Wastewater Technologies (AWT) Program, was developed to identify and evaluate technologies that can reduce environmental impacts at the wastewater treatment plants as identified by citizen groups. These include the space requirements, or the "footprint," of the treatment facilities, odor emissions, air emissions, the volume of sludge produced, and the truck traffic in the community. King County is interested in technologies that can increase the solids destruction efficiencies and/or increase the cake solids dryness from dewatering processes to reduce the amount of biosolids hauled off-site. King County currently operates two wastewater treatment facilities: the West Point Treatment Plant (WPTP) and the East Section Reclamation Plant (ESRP). The WPTP uses centrifuges to dewater sludge while the ESRP uses belt-filter presses.

Pilot plant testing of an autothermal aerobic digestion process, the VERTAD™ process (Vertical Aerobic Digester), was carried out at the ESRP. The purpose of the VERTAD™ Demonstration Project was to evaluate its potential for reducing the volume of biosolids while producing Class A biosolids with a minimal amount of odor. One possible mode of application for the VERTAD™ process is to operate the VERTAD™ at a short solids retention time (SRT) of 4 days before mesophilic anaerobic digestion. A laboratory study was conducted in conjunction with the VERTAD™ pilot plant operation to determine the effect of VERTAD™ pre-treatment on mesophilic anaerobic digestion performance, including volatile solids (VS) destruction and the dewaterability of the final solids.

Three mesophilic digester systems were operated at bench-scale to observe the effect of VERTAD™ pre-treatment. The systems were anaerobic digestion alone, fed thickened primary and waste activated sludge, and anaerobic digesters at two different SRTs fed the VERTAD™ product sludge. The main parameters used to evaluate digester performance include volatile solids (VS) destruction, gas production and its percent methane, and solids dewaterability. A capillary suction time test was used to test the dewaterability of the sludge by determining how easily water drains from the sludge and the effect of polymer dose.

Background

Anaerobic Sludge Digestion

Anaerobic digestion is a commonly used sludge stabilization process. Conventional "high rate" anaerobic digesters are mixed and heated with operation at mesophilic conditions (35°C) with SRTs ranging from 15-20 days, and with the hydraulic retention time (HRT) equal to the SRT. Anaerobic digestion can decrease the final mass of solids using less energy than aerobic systems since aeration is not required. Another benefit of anaerobic digestion is the production of methane gas, which can be collected and combusted to produce the energy needed to heat the digesters. Long SRTs are needed to stabilize the sludge and allow for the hydrolysis of particulate organic matter. Anaerobic systems are more sensitive to low pH values, feed variations, and some chemical solutions and can be more complicated to operate than aerobic digesters. According to the WEF Design of Municipal Wastewater Treatment Plants (1998), anaerobic digestion can yield 20% to 25% cake solids from a belt filter press using 3 to 8 g dry polymer/kg dry solids (6 to 16 lb dry polymer/ton dry solids).

Autothermal Aerobic Digestion Processes

Autothermal aerobic digestion (ATAD) uses the heat generated from the biological oxidation of organic matter to elevate the digester temperature between 45°C to 65°C without using external heaters. Energy is needed, however, to aerate the digester. The higher temperatures promote increased pathogen inactivation. The solids must be concentrated to at least 3% solids in order to minimize the energy used to heat water (WEF, 1998). ATAD SRTs can be as low as 5-6 days to meet Class A biosolids requirements, resulting in operating volumes significantly smaller than anaerobic digesters. ATADs have been reported to have severe foaming problems due to their higher loading rates and thermophilic temperatures, and generally have higher ammonia concentrations than other aerobic digesters (Grady, et al., 1998). Class A biosolids can be generated in ATADs by meeting certain temperature time conditions as shown in Figure 1 from the 503 biosolids regulations for the Time/Temperature requirements (provided by E&A Environmental Consultants, Inc., 1999).

The VERTAD™ reactor is an underground variation of the autothermal thermophilic aerobic digestion (ATAD) process. A VERTAD™ pilot unit was installed at the ESRP in 1998 and consists of a 20-inch diameter pipe extending 350 feet underground (Figure 2). The reactor was designed with a processing capacity of 500 to 2000 lbs of solids per day. During the study reported here, the VERTAD™ pilot plant reactor was operated at a 4-day SRT to digest thickened primary and secondary waste activated sludge.

In earlier tests, the VERTAD™ process was operated at SRTs as low as 3 to 4 days and achieved 43% VS reduction. The oxygen transfer efficiencies were about 50% based on off gas oxygen measurements (Brereton, 2000). Polymer vendors also performed bench-scale dewatering tests using a method that they felt could match the cake solids possible from a belt filter press. In these tests, a cake solids of approximately 30% was obtained for the VERTAD™ product, which is much higher than the 20% cake solids presently obtained from the ESRP anaerobic digester sludge using a dry cationic polymer (E&A Environmental Consultants, Inc., 1999).

There are essentially two main sections of the VERTAD™ process digester. The upper zone, from the surface to a 160-foot depth, is a complete mix zone. In this zone circulation of the sludge, through a 10-inch diameter tube (downcomer) inside the main shaft, is induced by an airlift that is created by injecting air into the reactor at the 160-foot depth. The pressure gradient created between the downcomer and the annular space causes the rapid mixing. The lower section is designed as plug flow with the discharge at the very bottom. The aeration for this section is near the bottom of the shaft at 315-foot depth. There is some mixing in this zone due to the aeration. The final zone, from 315 to 350 feet, is not aerated and is called a "soak" zone designed to utilize the residual oxygen from the lower zone as the solids undergo final degradation and pathogen kill (E&A Environmental Consultants, Inc., 1999). The thermophilic conditions in the VERTAD™ Process produces Class A biosolids with a minimal footprint or area requirement. With the present operation of a batch feed cycle every 4 hours, the VERTAD™ reactor must be operated at 64°C to meet Class A treatment, according to Figure 1.

Dual Digestion

Autothermal thermophilic digestion followed by anaerobic digestion has been termed dual digestion, and is a sludge stabilization alternative that can produce Class A biosolids. The additional contact time that an ATAD provides before anaerobic digestion may increase the amount of volatile solids destruction above that by the digester alone. However the high temperature pretreatment may not increase the rate of solids destruction in anaerobic digestion. Research by Ward (1997) showed that a 1-day ATAD pre-treatment did not increase the solids destruction rate in a subsequent anaerobic digester. If a conventional anaerobic digester with a 15-day SRT follows a 4-day SRT VERTAD™ process, the total volatile solids destruction efficiency should increase compared to a 15-day SRT digestion alone. This also suggests that the same VS destruction efficiency achieved with the 15-day SRT mesophilic anaerobic digestion alone may be obtained with a shorter SRT in an anaerobic digester following VERTAD™ treatment. Excess heat captured from the ATAD can also be used to help heat the anaerobic digester.

Dewatering Digested Sludge

The dewaterability of digester sludge has become an important concern for King County because of the issues related to hauling biosolids. The percent cake solids achieved from dewatering has a direct and significant impact on the haul cost. An increase from 20% cake solids to 22% can decrease the volume of solids for transport by almost 10%. Conditioners, usually organic polymers, are added to the sludge to enhance floc formation to separate free water in dewatering equipment. The chemistry of the polymer used can affect the floc characteristics and the final cake solids concentration

The dewaterability of a sludge depends on many factors including the solids concentration, the particle size, the amount of bound water, and the temperature and viscosity of the sludge. The soluble concentrations of lipids, proteins, and polysaccharides have also been claimed to play an important role in the dewaterability of the sludge (Novak et al., 1999, Chen et al., 1996). The specific polymer used for conditioning is specifically chosen to meet the properties of the sludge product as well as the method of dewatering. Filter presses and centrifuges exert different types of forces on the floc that can affect polymer choice. For example sludge dewatered on a filter press might need to form a floc with greater shear strength than one dewatered by centrifugation. The ESRP currently uses Perclor 7503 (CIBA) a cationic dry polymer in a 0.25% solution. The conditioned sludge is then placed on a belt filter press to produce 17-20% cake solids at an approximate polymer dose of 25 to 28 lb dry polymer /dry ton solids.

Many bench-scale methods have been used to evaluate the dewatering characteristics of different sludges. The specific resistance to filtration (SRF) was the first widespread method used to characterize sludge dewaterability. The SRF is a measure of the resistance to flow of filtrate through a porous media according to Darcy's law (Vesilind, 1988). The porous media is the filter cake formed having a unit weight of dry solids per unit filter area. This test uses a Buchner funnel to extract the water as the solids form into a cake under either vacuum or positive pressure up to 5 psi. The resistance is dependent on the solids concentration and can be affected by the amount of colloidal material (0.10 to 10 μm diameter) it contains (Karr and Keinath, 1978). Fine particles can blind the water passages and result in higher resistance measurements. The SRF is the slope of the line between the time per volume of filtrate versus the volume of filtrate alone. 250 mL is the typical sample volume used. This minimum volume requirement can limit the number of duplicate tests and polymer dosage tests run during bench scale tests.

The Capillary Suction Time test (CST) provides a quantitative measure of the ability of water to drain from a sludge. The test device has an electric timer that automatically records the time required for the filtrate to travel 1 cm through the filter paper. The absorbent filter paper applies a capillary suction onto the sludge and pulls the water out in a concentric ring (Figure 3). Each CST test requires approximately 7 mL of sample volume. Although this is purely an empirical test, the CST can be used to describe a factor called the filterability accurately (Vesilind, 1988). The filterability of a sludge is a constant that can be compared to different sludges. No direct correlation has been shown between the CST of a sludge and its performance on a belt filter press or a centrifuge. The CST was chosen to test the dewaterability of the sludge in this research to provide an indicator of changes in digester solids characteristics and due to the small volumes required for each test. With limited amounts of sludge available from the lab digesters, multiple polymer doses could be investigated using smaller volumes of the digester sludge.

Methods and Materials

Bench-scale Digesters

Three constantly mixed mesophilic anaerobic digesters were operated at 35°C in a constant temperature chamber. The control digester was fed the thickened waste primary/secondary sludge from the ESRP and operated at a 15-day solids retention time. At the end of the lab digester operation the digesters were opened and the final volume in the control digester showed that the actual operating SRT was only 11 days. Liquid had been lost due to earlier foaming problems, but the amount lost could not be determined until the digester was opened. Thus the control digester, without VERTAD pretreatment, will be referred to hereafter as an 11-day SRT digester. The other two digesters were fed 4-day SRT VERTAD™ product, and were operated at 15-day and 11-day SRTs respectively. The control digester and the 15 day VERTAD™ fed digester were in operation since June 18, 1999 while the 11 day VERTAD™ fed digester was started on July 5, 1999. VERTAD™ solids feeding to the laboratory digesters began on July 28, 1999.

The digester vessels were 4-liter glass reactors with 3 liters of liquid. The bottles had two ports near the bottom for sampling and feeding (Figure 4). The top was sealed with a rubber stopper containing a tygon tube that was connected to a Wet Flow Gas Meter (Precision Scientific Model 62627, Varlan Instruments, Chicago, IL) to provide continuous total gas production measurements. The gas was sampled off a port on the collection line and tested for methane concentration. Each digester was mixed using a 3-inch magnetic stir bar and a large Cimarron magnetic stir plate.

The digesters were fed daily using a wide-mouth 60-mL syringe. The feed was collected twice a week from the ESRP and stored in a 4°C room. Before the digesters were fed, an equal volume was removed from the digester. The sampling and feeding occurred within the same three hour time period every day (10:30 am to 1:30 pm). The digesters were brought to steady-state, defined as three SRTs of operation, or 45 days and 33 days respectively. All three digesters were operated under steady-state conditions from October 11, 1999 to November 12, 1999.

The control digester experienced a significant amount of foaming and scum formation throughout the duration of the lab operation. At one point before October, excess foam blocked the gas line and caused a pressure build-up which caused a loss in reactor volume. The estimated volume after the blow-out was 2.0 liters, but the actual volume was later found to be only 1.5 liters. The digester was fed and wasted assuming a 2.0-liter reactor volume to maintain a 15-day SRT, but the actual SRT of the control digester was closer to 11 days. Foaming was not a significant problem in either of the VERTAD™ fed digesters.

Performance Parameters

The digester pH, gas production and methane percentage were measured daily to monitor the digester operation. The steady-state analyses are described in Table 1 with most parameters measured three times a week for 5 weeks. In addition, the digester soluble polysaccharide and soluble protein were determined three times during steady state. The digester TKN and soluble COD were measured once during steady state. The CST dewatering test was done for each digester sludge at steady state. The tests were first run twice without polymer. Different polymer doses were then tested to determine the minimal CST values and optimal polymer dosage. The CST tests for a digester were all done on the same day.

Table 1 – Steady State Analyses		
Parameter	**Feed**	**Digester**
PH	D	D
Alkalinity	3x	3x
Gas production		D
% CH_4		D
TS	3x	3x
VS	3x	3x
Total COD	3x	3x
NH_4		3x

*D = daily measurements; 3x = 3 times/week

Dewatering Tests

CST measurements were done at room temperature (23°C ±2°C) with and without sludge conditioning with cationic polymer. Various polymer doses were added to the sludge samples in order to determine an optimum polymer dosage, which was the lowest amount of polymer needed to obtain a lower CST value which changed only modestly with higher polymer additions. A dry organic polymer Perclor 7503, which is used at the ESRP, was used to condition the sludge. A 0.50% polymer solution was made by mixing 1.0g of dry polymer into 200 mL of deionized water on a magnetic stir plate. After the dry polymer was thoroughly mixed into solution, the solution was allowed to cure by holding without mixing for 30 minutes before use.

The following equation describes how the polymer dosage in lb polymer / dry ton was determined:

$$\frac{lb\ \ active\ \ polymer}{dry\ \ ton\ \ solids} = \frac{(X)\ \left(\frac{0.5000g\ \ drypolymer}{100mL\ \ DI\ \ water}\right)\ \left(\frac{lb}{g}\right)}{(Y)\ \ (45mL\ \ sample)\ \left(\frac{1L}{1000mL}\right)\ \left(\frac{lb}{g}\right)\ \left(\frac{ton}{2000lb}\right)},$$

where

X = mL of polymer used,

Y = sludge in grams of dry solids/Liter, assuming specific gravity = 1.0

The polymer was added to 45 mL of sample, which was mixed on a small Cimarron magnetic stir plate with a 5/16 x ½" magnetic stir bar. The sludge was initially mixed at a low speed (level

3) to ensure that the sludge was homogenous. After the polymer was added to the stirring sample, the speed was increased to level 8 for all samples. This was the highest mixing intensity that could be sustained using the current stir bar size without the magnet becoming disengaged due to floc accumulation on the magnet. The high speed mixing time was set at 45 seconds for the control digester and 3 minutes for the VERTAD™ fed digesters. The mixing times were chosen based on visual observations of the floc at various mixing speeds and times to select conditions that resulted in a stable floc structure. A CST was run for each mixing combination to examine the effect of longer mixing times or increased intensity. Higher CSTs were expected when shearing of the floc occurs.

Shearing was determined from visual observations during mixing. The optimal times result in a very clear centrate and well-defined floc structure. Above those times, the floc was either broken up through shearing and deteriorated rapidly, or no difference in the floc structure was seen. The control digester experienced floc shear at mixing times above 15 seconds. The 15-day VERTAD™ fed and the 11-day VERTAD™ fed digester sludge needed a longer mixing time to obtain good floc structure, and the sludge did not seem to be affected by longer mixing times.

The optimal polymer dose curves of CST versus polymer dose were generated on different days. The polymer was made fresh daily to ensure consistent polymer quality. Each curve had a minimum of six different polymer doses, including a point with no polymer. At least duplicate tests were run at a particular polymer dose. Most samples were tested in triplicate or higher. The sludge samples used for a polymer dosage curve were stored in a 4°C cold room, up to 10 days, until enough volume was available for testing. The samples were brought up to room temperature (23°C ±2°C) before being conditioned and tested on the CST.

Analytical methods

pH and Alkalinity

The pH of the samples was measured with a Corning pH flat-surface combination pH electrode probe (Corning Inc., Corning, NY) and a Beckman Φ 220 pH meter (Beckman Instruments, Inc., Fullerton, CA). The same probe was used to measure the pH in titrations to determine the alkalinity of the samples and the feed according to Standard Methods procedure 2320B. Following this method, the alkalinity was calculated using the following equation

$$Alkalinity\ ,mgCaCO_3 / L = \frac{A \ x \ N \ x \ 50000}{mL \ \ sample}$$

where:

A = mL of acid used in the titration and

N = Normality of acid used, 0.02N H_2SO_4 for all analyses.

Total and Volatile Solids

Total and volatile solids measurements followed Standard Methods procedure 2540G. Sample volumes, approximately 5 to 15 mLs, were poured into pre-dried crucibles. The total solids samples were dried in the 103°C oven for over 24 hours. The samples were weighed for total solids after they had cooled in a desiccator to room temperature. The volatile solids concentrations were determined from weighing the samples after ignition in the 550°C muffler furnace oven for 30 minutes and cooled to room temperature in a desiccator.

Ammonia and Total Kjeldahl Nitrogen

An ORION Model 95-12 Ammonia ion-selective probe (Orion Research, Inc., Beverly, MA) was used to measure the ammonia nitrogen concentrations in the samples. The procedure followed Standard Method 4500-NH_3 D. The pH of the diluted samples was adjusted to 11 by adding Ionic Strength Adjuster (ISA) fluid, which contains 5M NaOH, 0.05M disodium EDTA, and 10% methanol. The samples were stirred on a magnetic stirrer while the measurements were being taken. The ammonia concentration was calculated from an internal standard curve on the Beckman Φ 220 pH meter (Beckman Instruments, Inc., Fullerton, CA). The standard curve was calibrated each time using 10 mg/L and 100 mg/L NH_4-N standard solutions.

The Total Kjeldahl Nitrogen (TKN) digestion was performed with the HACH Digestahl Digestion Apparatus (HACH Company, Loveland, CO). The digestion results in nitrogen concentrations that can be measured as ammonia with the Orion Model 95-12 Ammonia ion-selective probe. The aqueous liquid procedure in the HACH Digestahl® Digestion Apparatus Manual (1997) was used to digest the sludge samples.

Chemical Oxygen Demand

The HACH High Range (1-1500 mg COD/L) digestion reagent vials (HACH Company, Loveland, TX) were used to analyze the total chemical oxygen demand (COD) of the samples and the feed. This procedure is modified from the Standard Methods 5220D – Closed Reflux, Colorimetric Method where two mL of homogenized, diluted sample is placed in a prepared COD reagent vial. The vials are then digested on a heating block at 150°C for two hours. After the vials are cooled, the absorbance is read at 620 nm on a HACH DR/4000U spectrophotometer. The samples are diluted so that each vial contains less than 1500 mg/L COD. Feed and digester samples were composited separately for three days for sample analysis. These composites were blended and then diluted in a volumetric flask. All samples were measured in triplicates.

Soluble Protein

Soluble protein was measured using the Bio-RAD *DC* Protein Assay kit (Bio-RAD Laboratories, California). This colorimetric method is a modification of the Lowry method using bovine serum albumin as the protein standard. This assay uses a reaction between the protein and copper in an alkaline media and a subsequent reduction of Folin reagent. The blue color development is mainly due to the amino acids tyrosine and tryptophan. The sludge samples must be diluted such that soluble protein levels fall between 0.2 to 1.4 mg/mL protein. 100 μL of a sludge sample is mixed with 400 μL an alkaline copper tartrate solution and vortexed. Four mL of a dilute Foline reagent is then added and vortexed immediately. The vials develop 90% of its maximum color within 15 minutes. The samples are read for their absorbance at 750 nm on the HACH spectrophotometer.

Total and Soluble Carbohydrate

Carbohydrates were measure using the Dubois phenol-sulfuric acid method with glucose as the standard (Newton, 1999). Two mL of diluted sample was used. One mL of a 5% phenol solution was added to each sample. Five mL of concentrated sulfuric acid was then added and the mixture was vortexed to ensure thorough mixing. Once the vials reached room temperature (the reaction between phenol and sulfuric acid is extremely exothermic), the absorbance of the samples are read on a HACH spectrometer at 485 nm.

Gas Flow Measurement

The total gas flow was measured using three Precision Scientific Wet Flow Gas Meters (Varlan Instruments, Chicago, Il). The gas from the digester displaces a high viscosity oil which rotates a drum inside the meter. The drum in turn rotates the gears attached to the needle that reads the gas passing through the flow meter. The accuracy of each meter was tested using a peristaltic pump and a manometer before being placed in service after each digester. Each meter measured within 5% of the actual volume fed.

Gas Composition

Daily gas composition measurements were made using a gas chromatograph with a thermal conductivity detector (GC-TCD). 0.1 mL of the gas sample was collected from the gas line using a Mininert valve on a 1-mL syringe. The sample was injected into a Carle Series 100 Autosystem gas chromatogram (Chandler Engineering, Tulsa, OK). Methane and carbon dioxide concentrations were determined using a Hayesep Q80/100 column (Supelco, Bellefont, PA) and a HP 3396 Series II Integrator (Hewlett Packard, Avondale, PA). This column was operated at 60°C using helium as the carrier gas. The standard curve for the gas composition was generated using known ratios of methane and carbon dioxide. The percent of methane in the gas was determined by calculating the proportion of the methane peak area to the total area of the methane and carbon dioxide peaks combined. This method assumes that the gas is composed entirely of methane and carbon dioxide.

Capillary Suction Time Tests

The Triton Type 304B Capillary Suction Time (Triton Electronics, Essex, England) device was used to obtain dewaterability data for the sludge samples. Standard Method 2710G was followed in operating the CST. The sludge samples were either poured or syringed into the small liquid reservoir for analysis. A syringe was used on conditioned samples in order to obtain a representative sample. Whatman No. 17 chromatography grade paper cut into 7x9 cm sheets were used with the grain parallel to the long side of the CST device. The rougher side of the filter paper was placed uppermost to improve the reliability of the CST (Triton Electronics, Ltd., 1998). The time needed for the water to travel a known distance is recorded by an electronic timer which has contact points in contact with the chromatography paper (Figure 3).

Results and Discussion

Digester Performance

In this research, the performance of the three laboratory digesters were compared to determine the effect of VERTAD™ pre-treatment on anaerobic digestion. A summary of performance parameter values is shown in Table 2. The pH values of the three digesters were all very similar. The 15-day VERTAD™ fed digester exhibited higher alkalinity and ammonia concentrations than the 11-day VERTAD™ fed digester. This suggests that the 15-day VERTAD™ fed digester was able to destroy more protein. The 11-day VERTAD digester and the control digester had similar ammonia values. The gas production of the control digester and the 11-day VERTAD™ fed digester were comparable, 4.4 L/day and 4.3 L/day respectively. The 15-day VERTAD™ reactor had a significantly lower gas production at 3.4 L/day, but was fed less feed volume than the other two digesters. The gas production in L/g VS destroyed will be compared later.

Table 3 compares the percent VS and TS removals across the different digesters and the dual digestion systems (Vertad fed system with 11-day SRT and Vertad-fed system with 15-day SRT). The Vertad pretreatment did not greatly reduce the amount of solids destruction in mesophilic digestion for the dual digestion system. The %VS destruction for the 15-day and 11-day VERTAD™ fed digesters was 49% and 46% respectively, or about 94% and 88% of the VS destruction of the control digester (52% VS destruction).

Table 2—Summary of Digester Performance (Oct. 11 to Nov. 12). Average values shown with standard deviations.

Performance Parameter	Number of samples	Feed		Digesters		
		Thickened Primary and WAS (THS)	VERTAD™ Product	Control	15 day VERTAD™	11 day VERTAD™
Digester SRT (days)	—	—	—	15	15	11
TS (%)	14	6.0 ±0.26	5.1 ±0.51	3.5 ±0.13	3.2 ±0.18	3.4 ±0.19
VS (%)	14	4.8 ±0.22	3.8 ±0.37	2.3 ±0.11	2.0 ±0.12	2.1 ±0.14
Total COD (g COD/L)	7	80.6 ±3.62	61.3 ±3.52	39.6 ±2.09	35.6 ±2.20	35.7 ±1.97
PH	31	6.1 ±0.14	8.1 ±0.23	7.5 ±0.08	7.4 ±0.06	7.4 ±0.06
Alkalinity (mg/L $CaCO_3$)	9	3150 ±320	6130 ±550	9210 ±690	9880 ±220	9560 ±160
Ammonia (as mg N/L)	2	1070 ±70	1590 ±290	2270 ±120	2530 ±160	2240 ±500
L gas/day	33	—	—	4.4 ±0.9	3.4 ±1.1	4.3 ±1.1
% CH_4	11	—	—	63 ±2	58 ±2	57 ±1
TKN (as mg N/L)	1	3350	3450	3640	3580	3280
Total Polysaccharides (mg glucose/L)	4	38500 ±4400	8770 ±870	5510 ±950	3910 ±410	4070 ±929

Table 3—Effect of VERTAD™ pre-treatment on VS reduction. Average values shown with standard deviations.

	% VS removal across digesters	% VS removal for combined system	% TS removal across digester	% TS removal for combined system
Control 11-day SRT	52 ±1.7	—	42	—
VERTAD™ fed 15 day SRT	49 ±0.9	66 ±0.5	37	55*
VERTAD™ fed 11 day SRT	45 ±1.2	63 ±0.8	33	52*

* Based on % VS removal reported for the VERTAD™ process and calculation of 29% TS removal for the VERTADTM process using same water loss as for the % VS removal.

Based on the average thickened solids concentrations fed to the VERTAD™ system and the VS and TS concentrations after VERTAD treatment only 20% VS destruction is calculated. However, King County reported that a significant water loss occurs in the VERTAD™ process so that the process effluent flow rate does not equal the influent flow rate. The water loss is due to the heat generation and saturated air leaving the reactor. Accounting for this water loss a VS destruction in the VERTAD™ reactor was reported to be 33% (σ=1.9) by King County. The two dual-digestion processes thus achieved 66% and 64% total VS reduction respectively. This is a 27% and 23% increase in overall solids destruction over the control digester, which had only an 11-day SRT compared to a total of 15 and 19 days of sludge digestion time for the dual digestion system. Based on data in Ward's thesis (1997) showing percent VS destruction as a function of SRT, the increased VS destruction efficiency for the dual digestion process (VERTAD™ plus anaerobic digestion) can be attributed to the 4-day sludge digestion time in the VERTAD™ process and is likely not due to increased anaerobic digestion rates following VERTAD™ treatment. A one-way Analysis of Variance (ANOVA) test shows that the increase in VS destruction efficiency is a statistical difference as seen in Appendix A.

Table 3 also compares the percent TS removal for the mesophilic digester alone and the dual digestion system. The dual digestion system had TS removal efficiencies of 55 and 52 percent, respectively, for the 15-day and 11-day VERTAD™ systems, compared to 42 percent for the 11-day control digester. The improved TS removal efficiency is due to the higher digester time in the VERTAD™ process and anaerobic digester. Comparing a VERTAD dual digestion system to an 11-day SRT mesophilic digester the solids reduction is about 20 percent.

Table 4 compares the gas production for the control and two VERTAD-fed digesters. The two VERTAD™ -fed digesters exhibited similar gas production rates normalized to VS destroyed. The L-gas per g-VS removed were 0.94 and 0.92 for the 15-day and 11-day digesters respectively. These values are within the range of normalized gas production reported of 0.8 to 1.1 L gas/g VS reduced (WEF, 1998). The accepted estimate of liters of CH_4 produced per g-COD removed is 0.35 for conventional anaerobic digestion at 35°C (WEF, 1998). The two

VERTAD digesters averaged 0.39 and 0.36 L-CH_4/c-COD removed. The control digester produced 1.8 L-gas/g-VS removed and 0.51 L-CH_4/g-COD removed, much higher than the reported values. These elevated values suggest an error in the gas production readings for the control digester.

Table 4—Gas Production and COD Removal in Lab Digesters. Average values shown with standard deviations.

	Control	**15-day VERTAD™**	**11-day VERTAD™**
SRT (days)	11*	15	11
L gas/day	4.4 ±0.9	3.4 ±1.1	4.3 ±1.1
Δ VS(g)/day	2.5	3.6	4.6
L gas/Δ VS(g)	1.8	0.94	0.92
% CH_4	63 ±2	58 ±2	57 ±1
L CH_4/day	2.8	2.0	2.5
g $COD_{removed}$/day	5.5	5.1	7.0
L CH_4/g COD_{rem}	0.51	0.39	0.36

Assuming equal methane production per gram of VS destroyed, the VERTAD™ pre-treatment would decrease the methane production by about 6% in the 15-day digester and 12% in the 11-day digester. The lower methane levels results in less potential energy for the treatment plants, but the decreased production rates may not affect the treatment plants adversely. Less methane would be needed to heat the feed since the VERTAD™ product is at least 60°C. The methane would only have to be used to maintain the mesophilic temperatures in the digester. On the negative side of an energy balance, the VERTAD™ process requires energy to aerate. An energy balance including the VERTAD™ process energy needs is necessary to evaluate the combined system.

Dewaterability Tests

Figures 5 through Figure 8 show the CST values as a function of the polymer dose for each of the digesters and the VERTAD™ product. The error bars represent the standard deviation derived from duplicate tests at each polymer dose. The optimum polymer dose is considered the polymer dose that caused the CST to drop dramatically to a lower plateau Above this dosage, only minimal changes in the CST are observed.

Table 5 summarizes the optimal polymer doses and the related CSTs. The initial CST of all three digester sludges are very similar between 2100 and 2400 seconds. The VERTAD™ product itself had a lower CST at 1480 seconds. The optimal polymer dose was much higher for the VERTAD™ product and the VERTAD™ fed digesters compared to the control digester. The dual-digestion solids required approximately three times the amount of polymer needed for the control anaerobic digestion solids (about 90 lb/ton versus 30 lb/dry ton respectively) to achieve CSTs lower than 40 seconds. For comparison, the ESRP currently conditions the plant's solids using 24 to 28 lb polymer/dry ton for the belt filter presses. Although the VERTAD™ fed digestion solids required more polymer to achieve optimal CSTs, visual observations of the solids at the opimtal polymer dose showed the VERTAD™ fed digester solids to have a tougher floc that may have a higher resistance to shearing..

Table 5—Comparison of CST test results for the control digester to the VERTAD™ fed digesters.

	Control	15 day VERTAD™	11 day VERTAD™	VERTAD™ product
Unconditioned CST (sec)	2240 ±8.8	2110 ±65	2410 ±100	1480 ±360
Minimal CST (sec)	40 ±5	22 ±17	20 ±5	20 ±11
Optimum Polymer Dose (lb/dry ton)	30	95	90	100
Optimum Polymer Dose (g/kg solids)	15	48	45	50

The CST data can not be used to predict the performance of sludge dewatering equipment. The optimum polymer dose determined in these tests can also be different in a full-scale operation and the differences may vary with the type of dewatering equipment. The CST is affected by the quality of the centrate since suspended particles can blind the filter paper. Tests with centrifugation do not require high quality centrates and the polymer needed may be decreased to achieve comparable cake solids since the centrate is recycled within the plant (Brereton, 2000).

The flocs formed by these dual-digested sludges were longer, stringier, and harder to break apart than the anaerobic digestion only products which were smaller and appeared more granular. The floc formed by the VERTAD™ product was very strong and required cutting the floc to get a small enough sample into the CST thimble. The color of the sludge from both VERTAD™ fed digesters changed from the traditional black color to a brown, less viscous effluent after approximately 2 SRTs. The sludges seemed to have retained some of the visual qualities of the VERTAD™ product that the digesters were fed. The odor of the VERTAD™ fed sludge also seemed a little sharper and not as sweet as the odor from the control digester sludge. An odor panel was not assembled to characterize the difference or offensiveness of the new smell.

The percent cake solids can significantly affect the cost of the solids removal. Higher percent cake solids can decrease truck traffic through the community and decrease the overall costs of hauling the solids away. The economics of operation and removal are site specific, but the polymer costs for dewatering can be compared to the ultimate hauling costs. If the VERTAD™ fed digesters can significantly reduce the final sludge volume for hauling, a net total cost reduction is possible even with higher polymer requirements. A convenient bench-scale technique to assess the potential cake solids is needed.

Digestion Products

The soluble concentrations for COD, polysaccharides, and protein were analyzed to determine if the thermophilic process affected the final digestion products. Both VERTAD™ fed digesters contained over double the soluble protein and polysaccharides concentrations in solution than the control digester (Table 6). The two VERTAD™ fed digester sludges also exhibit similar soluble COD concentrations, approximately 1.5 times the control digester concentrations.

Table 6—Summary of soluble products. Average values shown with standard deviations.

	Control	15 day VERTAD™	11 day VERTAD™	VERTAD™ product	THS
Soluble COD (g COD/L)	2.00	3.61	3.89	4.95	4.89
Soluble Polysaccharide (mg glucose/L)	222±27	422±120	421±66	1440±260	325±46
Soluble Protein (mg protein as BSA/L)	400±19	1020±38	1080±82	2340±210	7210±190

Figures 9 and 10 plot the polymer demand of each of the digesters versus their soluble protein and soluble carbohydrate concentrations respectively. A trend of increased polymer demand with increased soluble protein and soluble polysaccharide is observed. This follows the trend of higher soluble protein concentrations and poorer dewaterability, based on CST tests, reported by Novak et al. (1999 WEFTEC papers) for both aerobically and anaerobically digested sludge. Novak et al. (1999) were also not able to characterize cake solids for the digested solids investigated.

Conclusions

Bench-scale anaerobic mesophilic digesters were operated to determine the effect of VERTAD™ pre-treatment on mesophilic digester performance and sludge characteristics. The control digester was operated at an 11-day SRT and fed thickened primary/waste activated sludge. The other two digesters were operated at 15-day and 11-day SRTs respectively and fed the VERTAD™ product. Throughout the research period, the VERTAD™ process was operated at a 4-day SRT.

The percent VS destruction efficiency was higher for the dual-digestion system at 66% and 64% for the 15-day and 11-day digesters respectively, compared to 52% destruction in the control digester. For percent TS destruction efficiency, the dual digestion system had 55% and 52%

removal efficiencies for the 15-day and 11-day digesters respectively, compared to 42% destruction in the control digester. The higher destruction efficiencies can be attributed to the longer digestion time provided by the VERTAD™ pretreatment and longer digestion time in the case of the 15-day VERTAD™ -fed digester. The pre-treatment thus results in a decrease in the final solids mass of about 20% compared to mesophilic digestion alone.

The solids destruction efficiency and gas production across the mesophilic digesters was not greatly decreased following the VERTAD™ pre-treatment. The VS destruction averaged 49% and 46% for the 15-day and 11-day digesters respectively, compared to 52% for the control digester. Since the methane production is expected to be proportional to the VS destruction, the two VERTAD™ fed digesters would expect to produce 6% and 12% less methane than the mesophilic digester alone.

No foaming was observed in the two VERTAD™ fed digesters while foaming posed a significant problem in the control digester.

The dewaterability of the digester contents was assessed with the Capillary Suction Time test. Similar CSTs were attained from each of the digesters, but the amount of polymer demand rose significantly for the dual-digested sludges. The polymer doses required to achieve a minimum CST value for both VERTAD™ fed sludges was nearly triple the polymer needed for the control digester. The control digester required about 30 lb polymer/dry ton which compares to the 24-28 lb/dry ton used at the ESRP for the belt filter presses.

The VERTAD™ fed digesters also had much higher soluble protein and soluble polysaccharide concentrations, approximately double the mesophilic digester's concentrations. The increased polymer demand seems to trend with an increase in soluble protein and polysaccharide concentrations. The mesophilic digestion following VERTAD™ treatment did not appear to improve dewaterability of the VERTAD™ sludge.

Future Research

This bench-scale research provided information on how the VERTAD™ process effected anaerobic digestion in a dual digestion process. For the short operating SRT used in the VERTAD™ process, there was not a significant impact on the digestion rates in the anaerobic process following VERTAD™ treatment.

The main issue that effects the benefits and economics of VERTAD™ pretreatment is how VERTAD™ treatment effects sludge dewaterability. The CST bench-scale testing were not adequate to assess the possible cake solids following belt filter press or centrifuge dewatering of the dual digestion sludge. Pilot plant digestion with pilot plant dewatering equipment is needed to better assess VERTAD™ treatment benefits. More research is needed to develop bench scale test methods to assess sludge dewaterability in a way that can relate bench scale results to possible cake solids from full scale dewatering equipment.

The VERTAD^{TM} process in this study used a 4-day SRT. Lower SRTs should be tested to determine if a more optimal dual digestion design can be developed. The effect of the VERTAD^{TM} SRT on sludge characteristics should be investigated to develop a more optimal dual digestion system design. This may best be done in pilot plant studies unless a valid bench scale procedure to assess dewaterability is developed.

Research is needed on pathogen destruction kinetics to provide more specific criteria for selection of the time and temperature of the pasteurization step in dual-digestion processes. Tests should be performed to determine the lowest VERTAD™ SRT that can be operated to maintain class A biosolids to decrease the overall detention time allowing the footprint to decrease.

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Appendix B—Volatile Solids Reduction Calculations. Average values shown with standard deviations.

	Control Digester	σ	15 day VERTAD	σ	11 day VERTAD	σ	VERTAD Product	σ
% VSR (UW)	51.54	1.73	48.95	0.88	45.63	1.20	15.99	6.73
Overall VSR (UW)	51.54	1.73	57.58	2.56	54.36	2.74	——	——
% VSR (ESRP)	56.93	2.66	48.01	3.86	44.68	4.75	33.20	1.90
Overall VSR (ESRP)	56.93	2.66	62.89	3.14	60.59	3.38	——	——
VS/TS ratio (ESRP)	53.81	4.69	61.99	3.51	59.90	3.71	——	——
VS/TS ratio (UW)	53.12	4.26	61.39	1.87	59.25	1.64	——	——
combined VSR	——	——	65.91	0.49	63.70	0.80	——	——

Volatile Solids Reduction Equations:

$$\% \text{ VS reduction} = \frac{VS_{in} * Q_{in} - VS_{out} * Q_{out}}{VS_{in} * Q_{in}} * 100$$

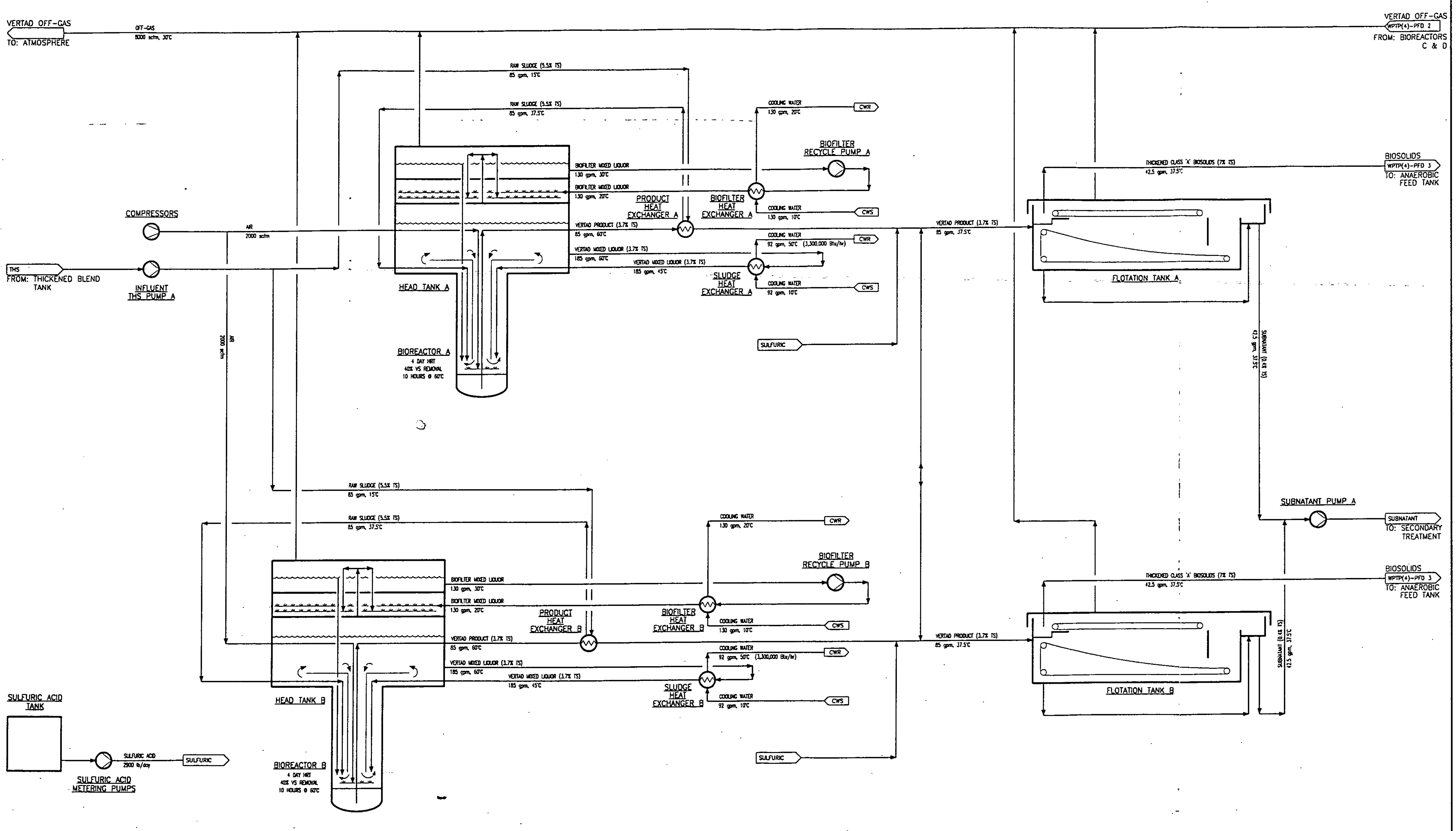
$$\text{VS/TS ratio} = \frac{v_{in} - v_{out}}{v_{in} - (v_{in} * v_{out})} * 100,$$

Where: $v = \text{VS/TS}$

$$\text{Combined VSR} = \left[\left(\frac{VSR_{VERTAD(tm)}}{100}\right) + \left(1 - \frac{VSR_{VERTAD(tm)}}{100}\right) * \left(\frac{VSR_{digester}}{100}\right) \right] * 100$$

APPENDIX B

Process Flow Diagrams



Process Flow Diagram for Alternative WP1
Figure 3-2 1 of 3

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADJUSTED FLOW VALUES	04JUL00	PR	JG		

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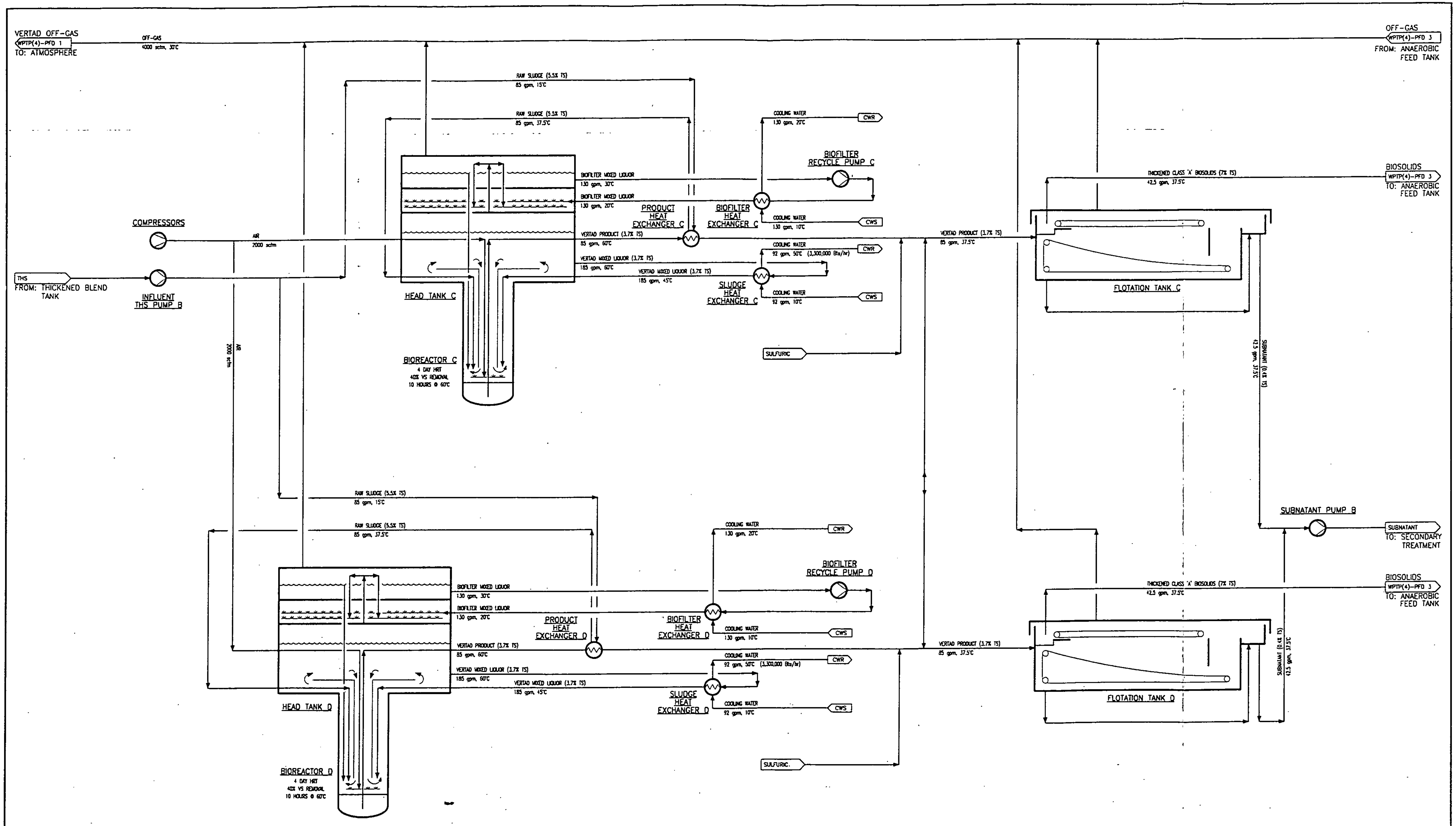
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN99	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
WESTPOINT TREATMENT PLANT
4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	WPTP(4)-PFD 1	B



Process Flow Diagram for Alternative WP1
Figure 3-2 2 of 3

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

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REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
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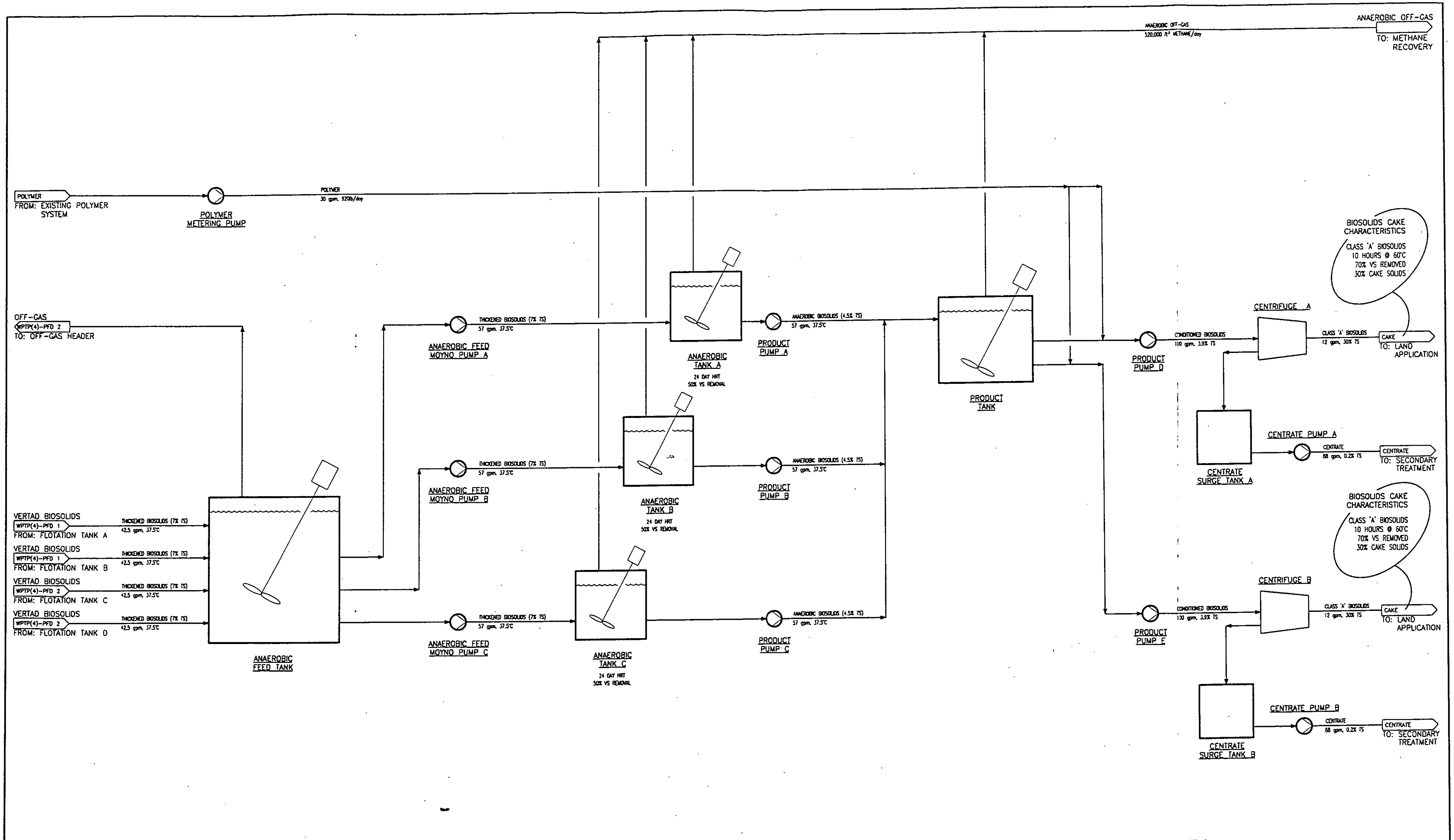
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
WESTPOINT TREATMENT PLANT
4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	WPTP(4)-PFD 2	B



Process Flow Diagram for Alternative WP1
Figure 3-2 3 of 3

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		

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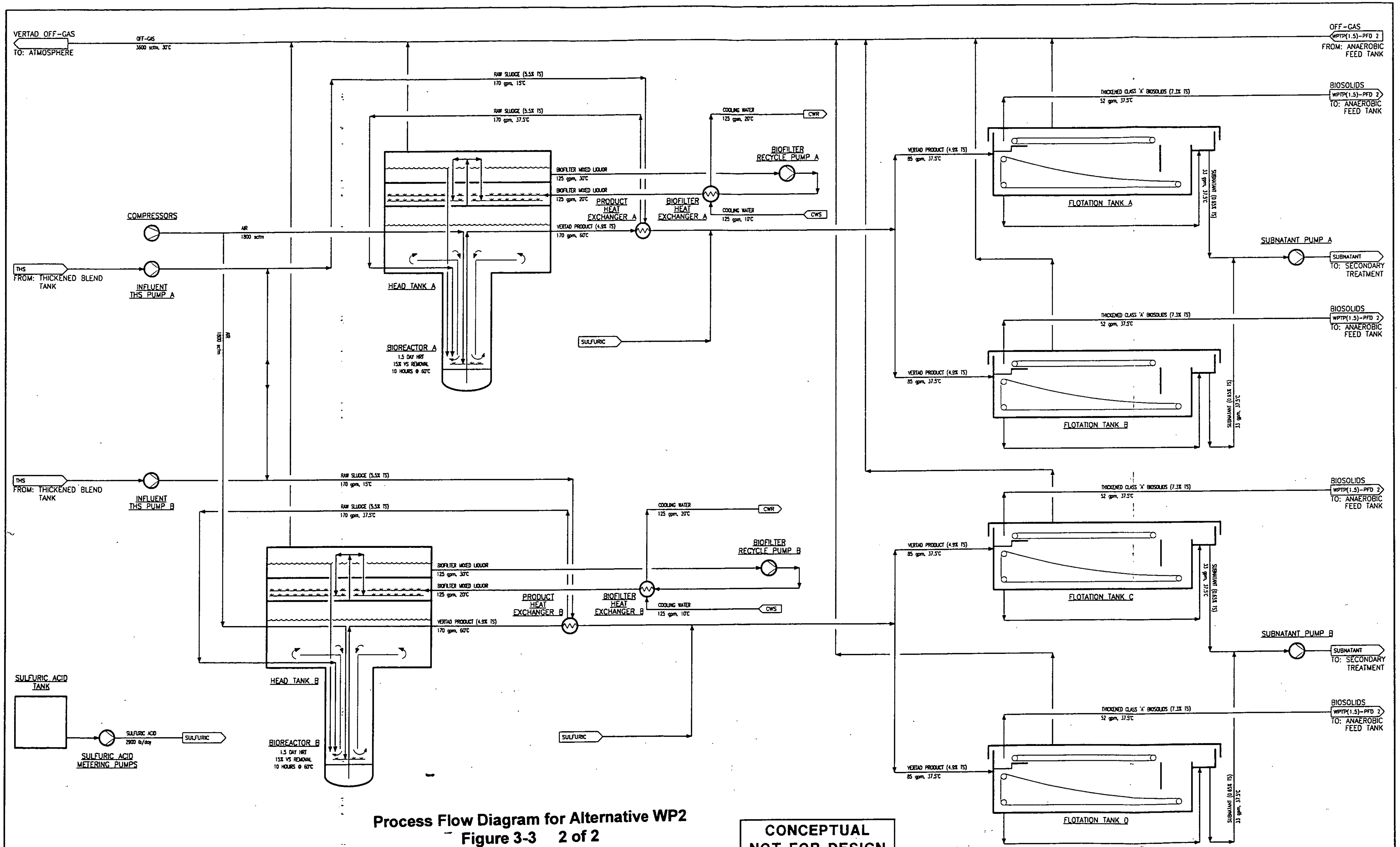
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN99	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
WESTPOINT TREATMENT PLANT
4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	WPTP(4)–PFD 3	A



Process Flow Diagram for Alternative WP2
Figure 3-3 2 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

○ PUMP
○ BLOWER
○ HEAT EXCHANGER
○ COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADJUSTED FLOW VALUES	04JUL00	PR	JG		

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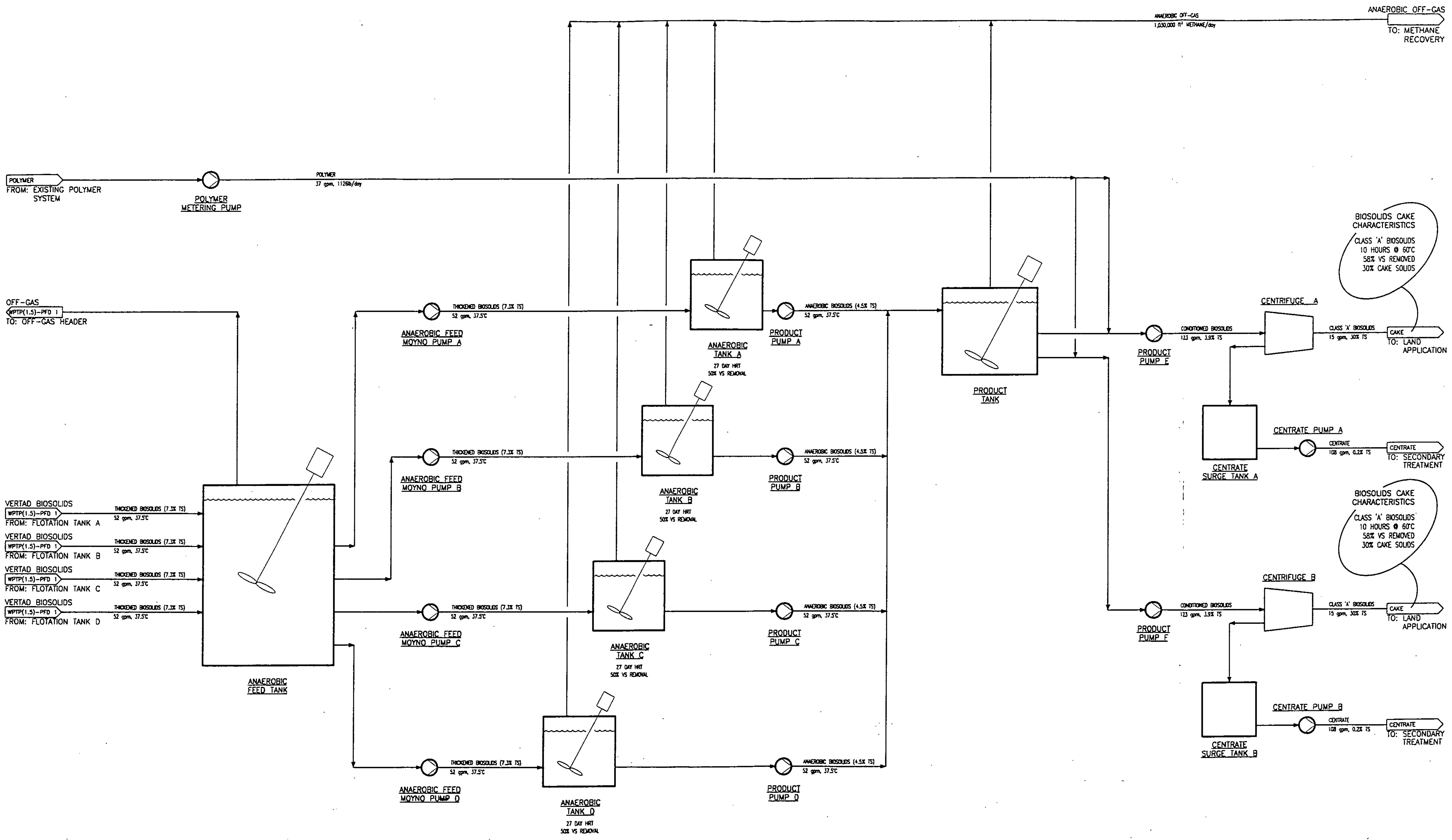
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN99	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
WESTPOINT TREATMENT PLANT
1.5 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	WPTP(1.5)-PFD 1	B



Process Flow Diagram for Alternative WP2
Figure 3-3 1 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		

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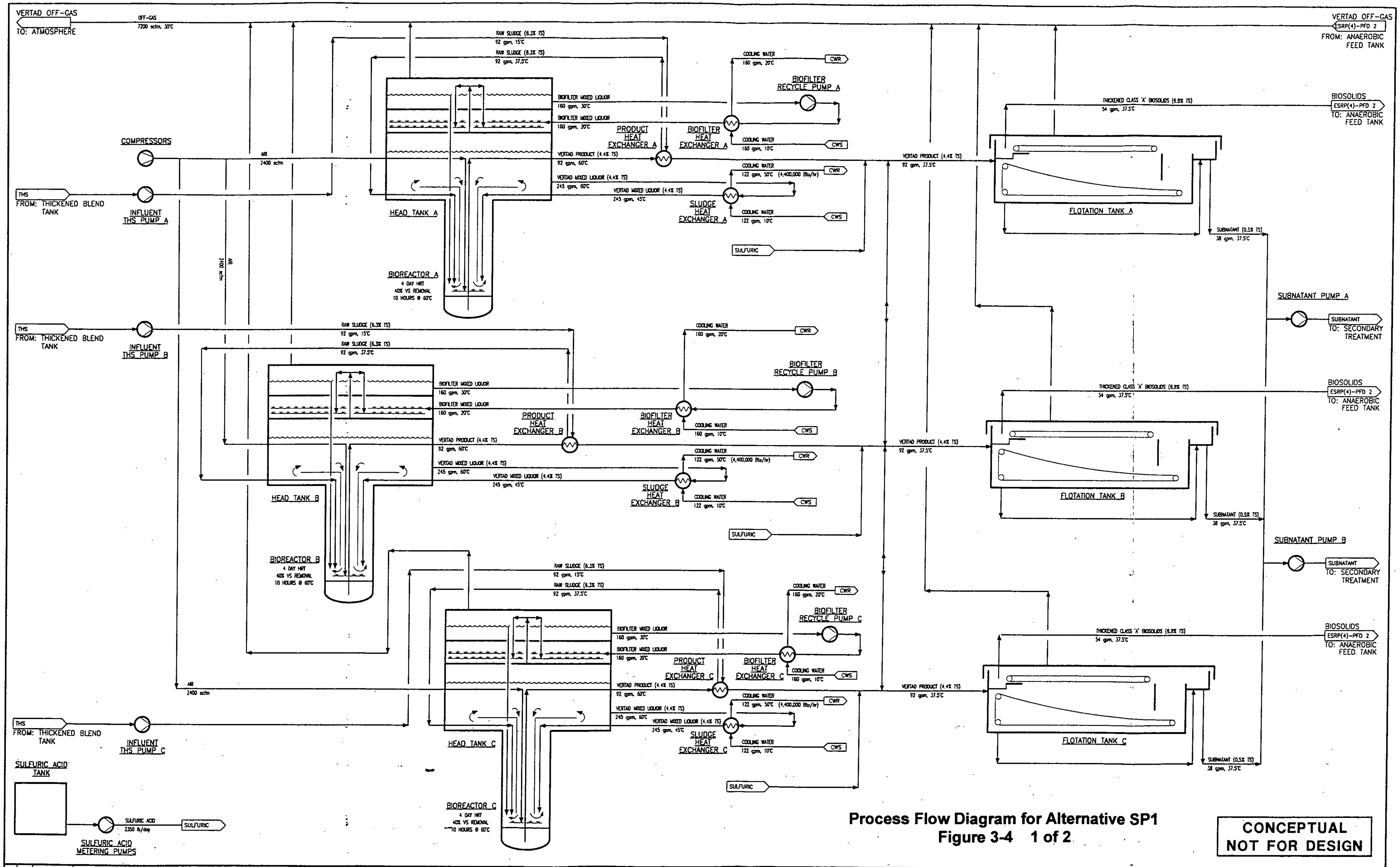
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
WESTPOINT TREATMENT PLANT
1.5 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	WPTP(1.5)-PFD 2	A



Process Flow Diagram for Alternative SP1
Figure 3-4 1 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADJUSTED FLOW VALUES	04JUL00	PR	JG		

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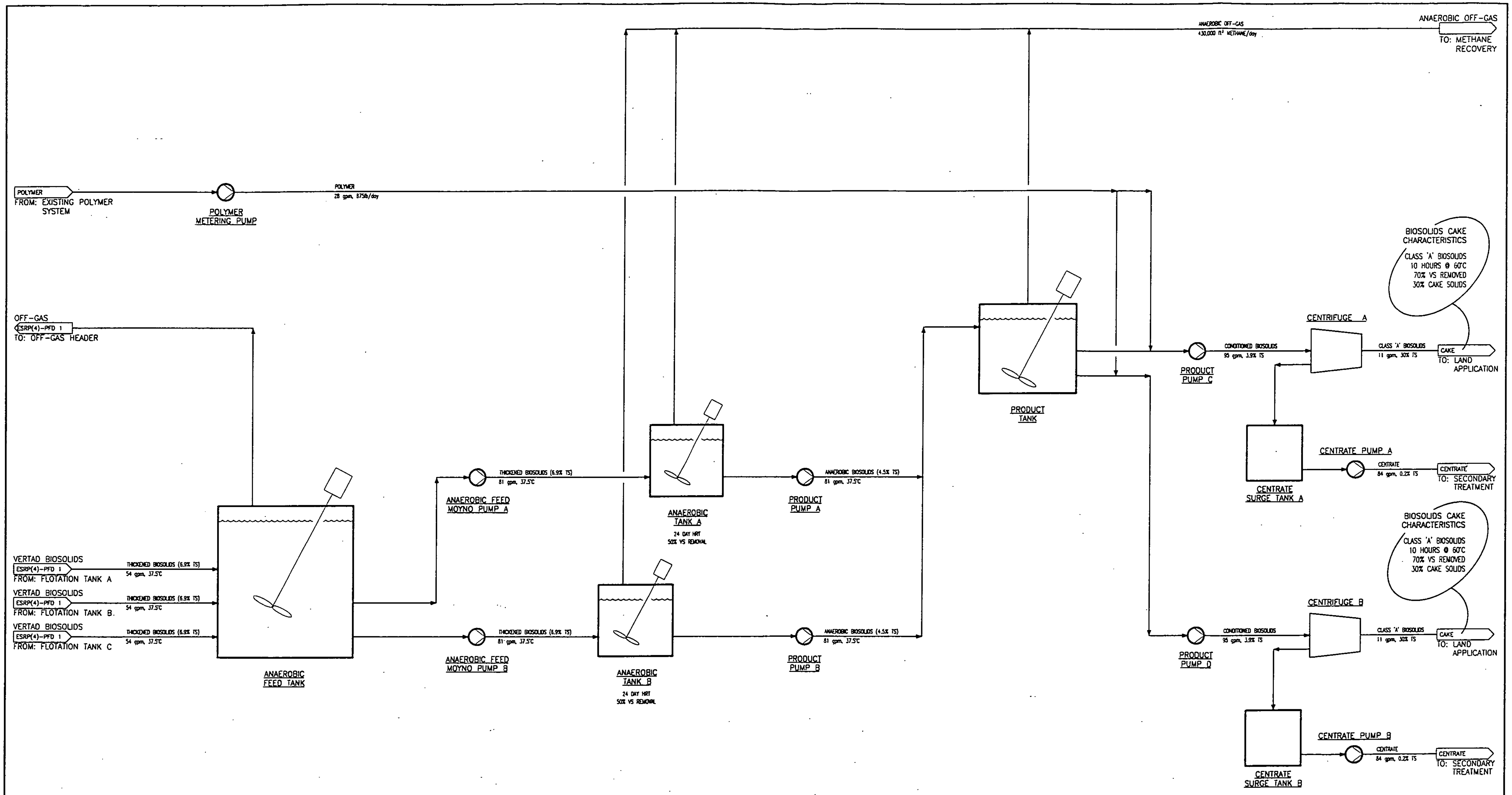
SCALE: NTS
PROJ MGR
DWG CHK
DRAWN 27APR00 PR
DES CHK
DESIGN 06JAN99 JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
EAST SECTION RECLAMATION PLANT
4 DAY HRT VERTAD

PROJECT NO
DRAWING NO ESRP(4)-PFD 1
REV B



Process Flow Diagram for Alternative SP1
Figure 3-4 2 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		

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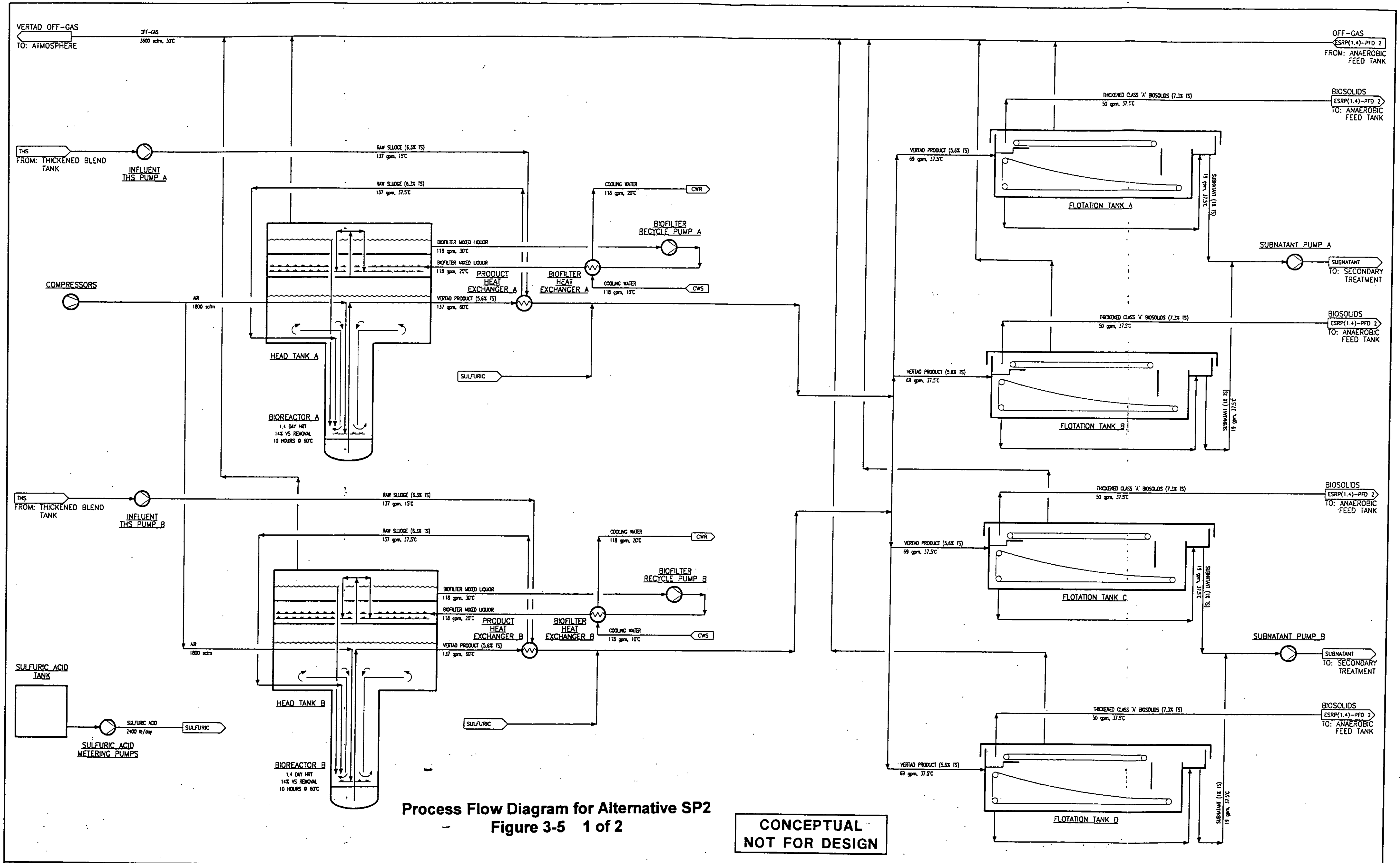
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	08JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
EAST SECTION RECLAMATION PLANT
4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	ESRP(4)-PFD 2	A



Process Flow Diagram for Alternative SP2
Figure 3-5 1 of 2

CONCEPTUAL NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

○ PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADDED REDUNDANCY	04JUL00	PR	JG		

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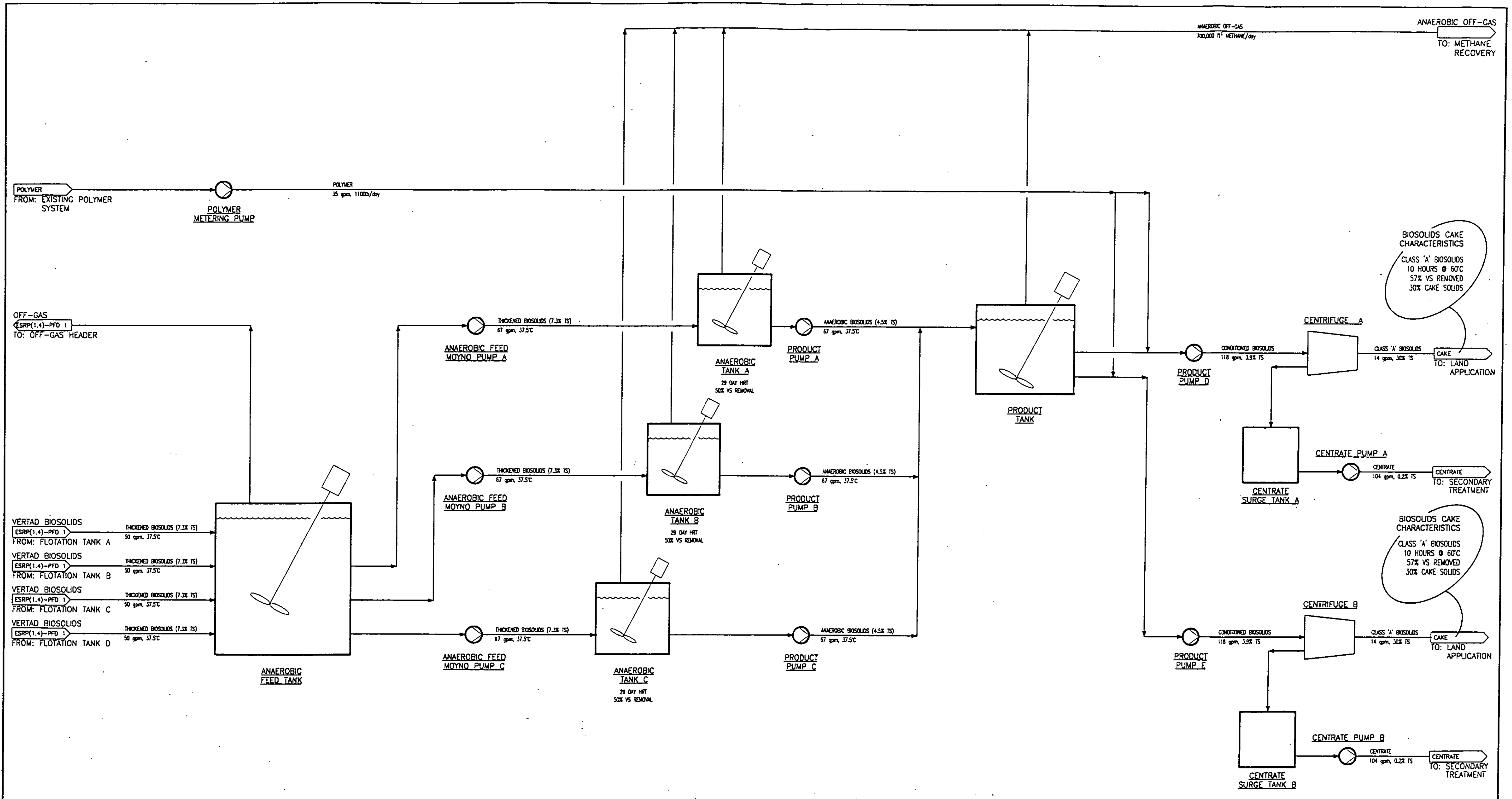
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	08JAN00	JG

KING COUNTY
VERTAD SYSTEM CONCEPTUAL DESIGN
EAST SECTION RECLAMATION PLANT
1.4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	ESRP(1.4)-PFD 1	B



Process Flow Diagram for Alternative SP2
Figure 3-5 2 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS NO	DWG NO	TITLE

NOTES:

- PUMP
- BLOWER
- HEAT EXCHANGER
- COMPRESSOR

DRAWING REVISIONS REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		

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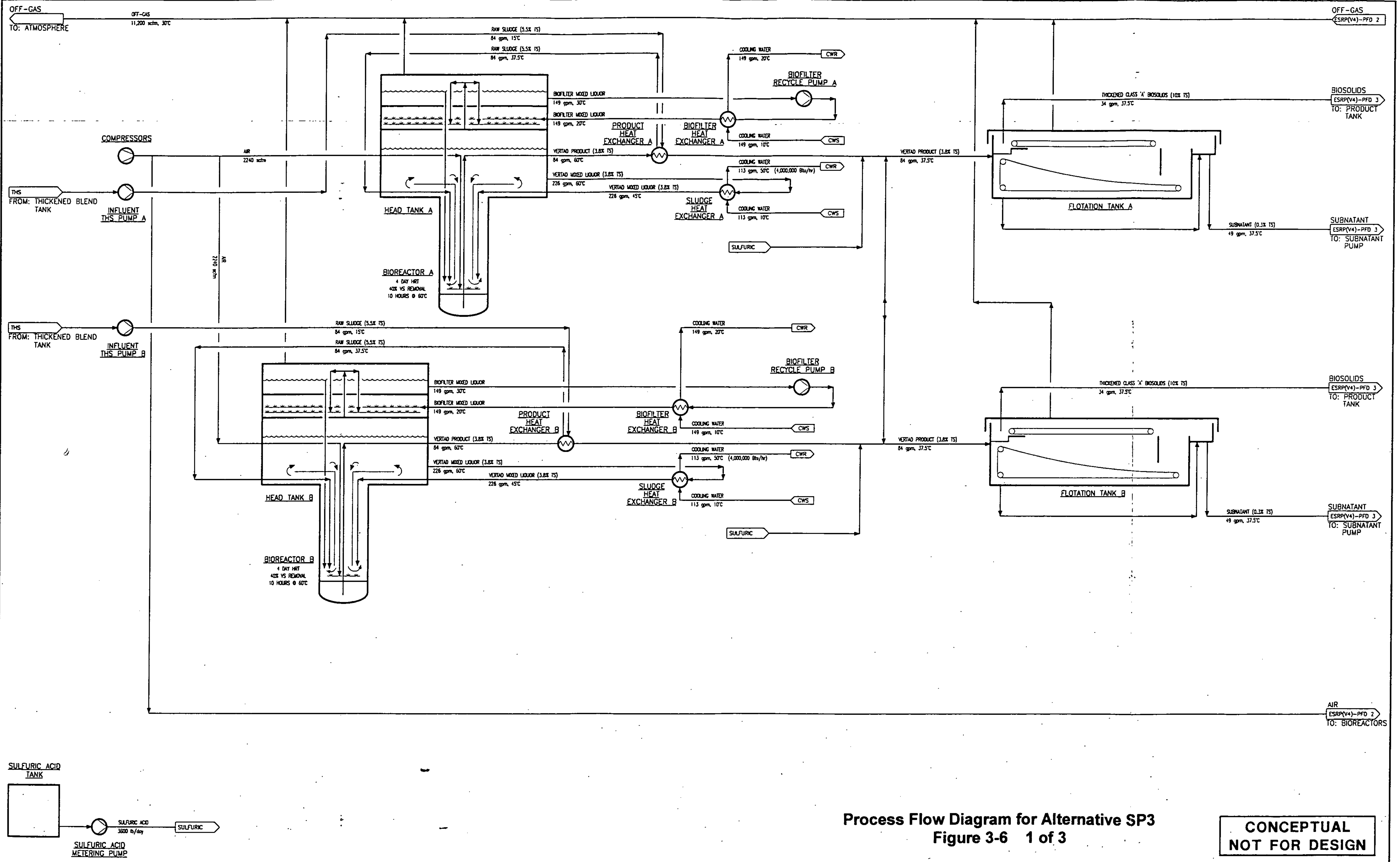
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	08JAN99	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
EAST SECTION RECLAMATION PLANT
1.4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	ESRP(1.4)-PFD 2	A



Process Flow Diagram for Alternative SP3
Figure 3-6 1 of 3

**CONCEPTUAL
NOT FOR DESIGN**

REFERENCE DRAWINGS NO	DWG NO	TITLE

NOTES:

- ○ PUMP
- ○ BLOWER
- ○ HEAT EXCHANGER
- ○ COMPRESSOR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	04JUL00	PR	JG		

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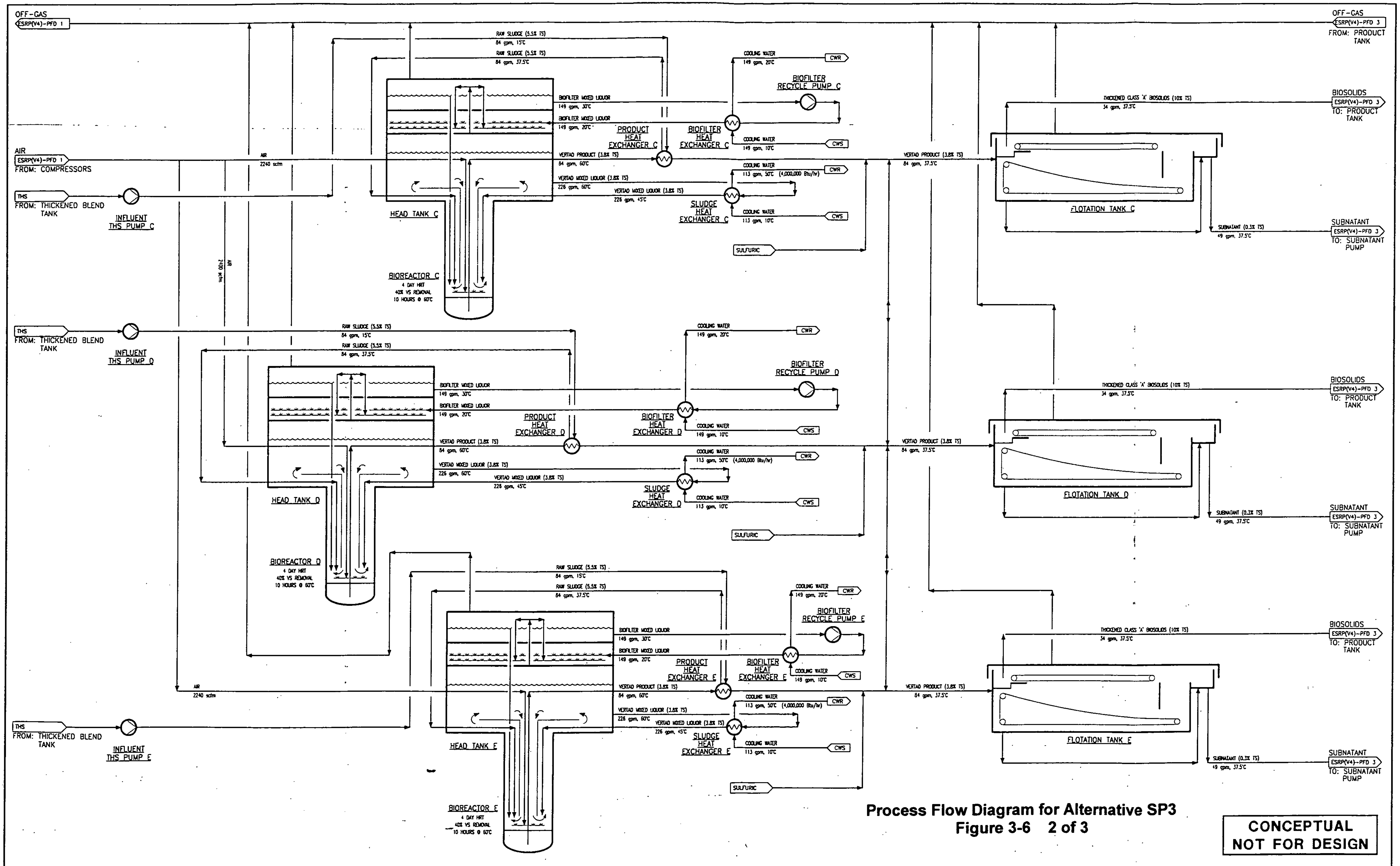
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	08JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
EAST SECTION RECLAMATION PLANT
PEAK WEEK–4 DAY HRT VERTAD ONLY

PROJECT NO	DRAWING NO	REV
	ESRP(V4)–PFD 1	A



Process Flow Diagram for Alternative SP3
Figure 3-6 2 of 3

CONCEPTUAL NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	04JUL00	PR	JG		

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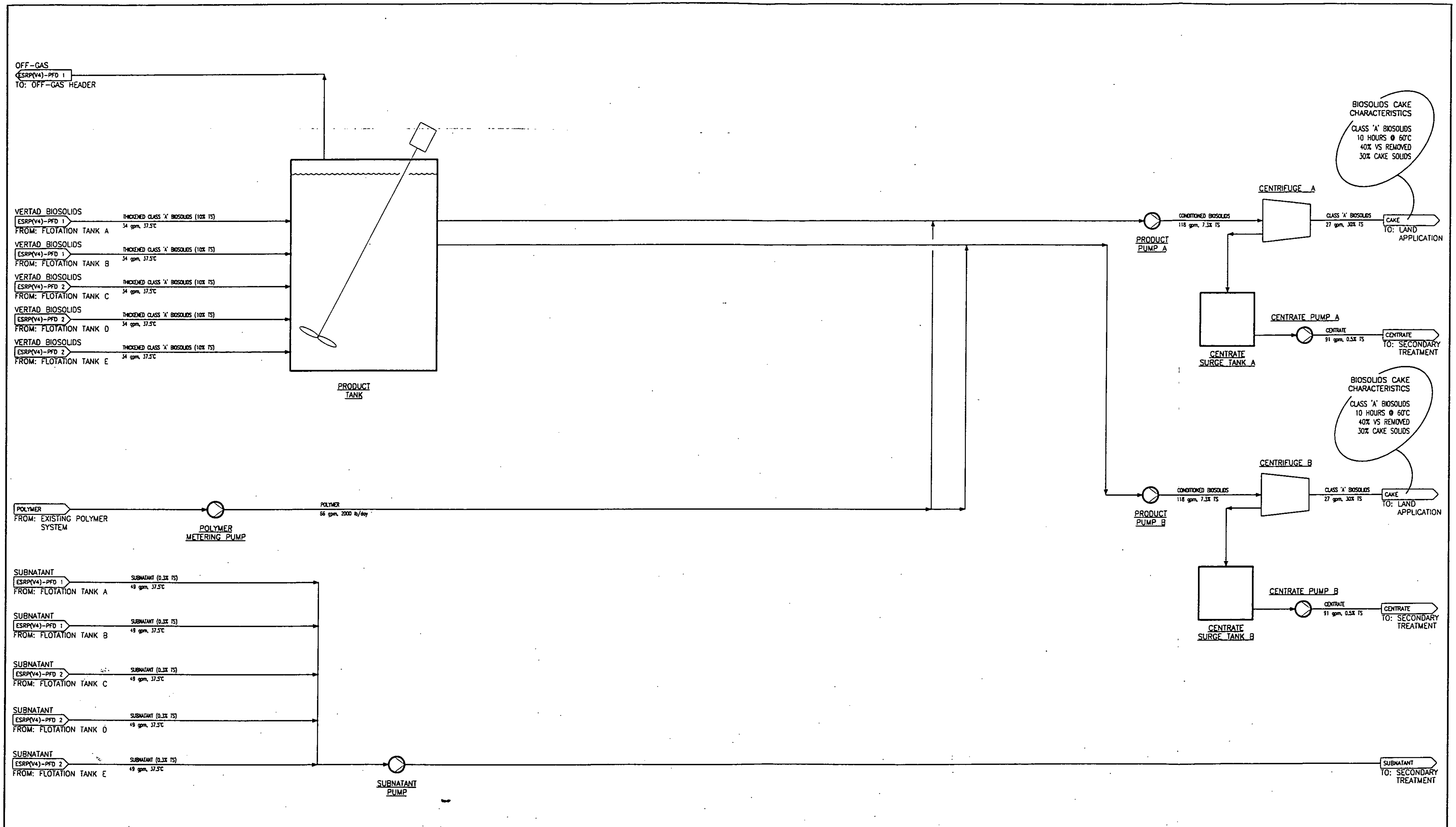
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN99	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
EAST SECTION RECLAMATION PLANT
PEAK WEEK-4 DAY HRT VERTAD ONLY

PROJECT NO	DRAWING NO	REV
	ESRP(V4)-PFD 2	A



Process Flow Diagram for Alternative SP3
Figure 3-6 3 of 3

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	04JUL00	PR	JG		

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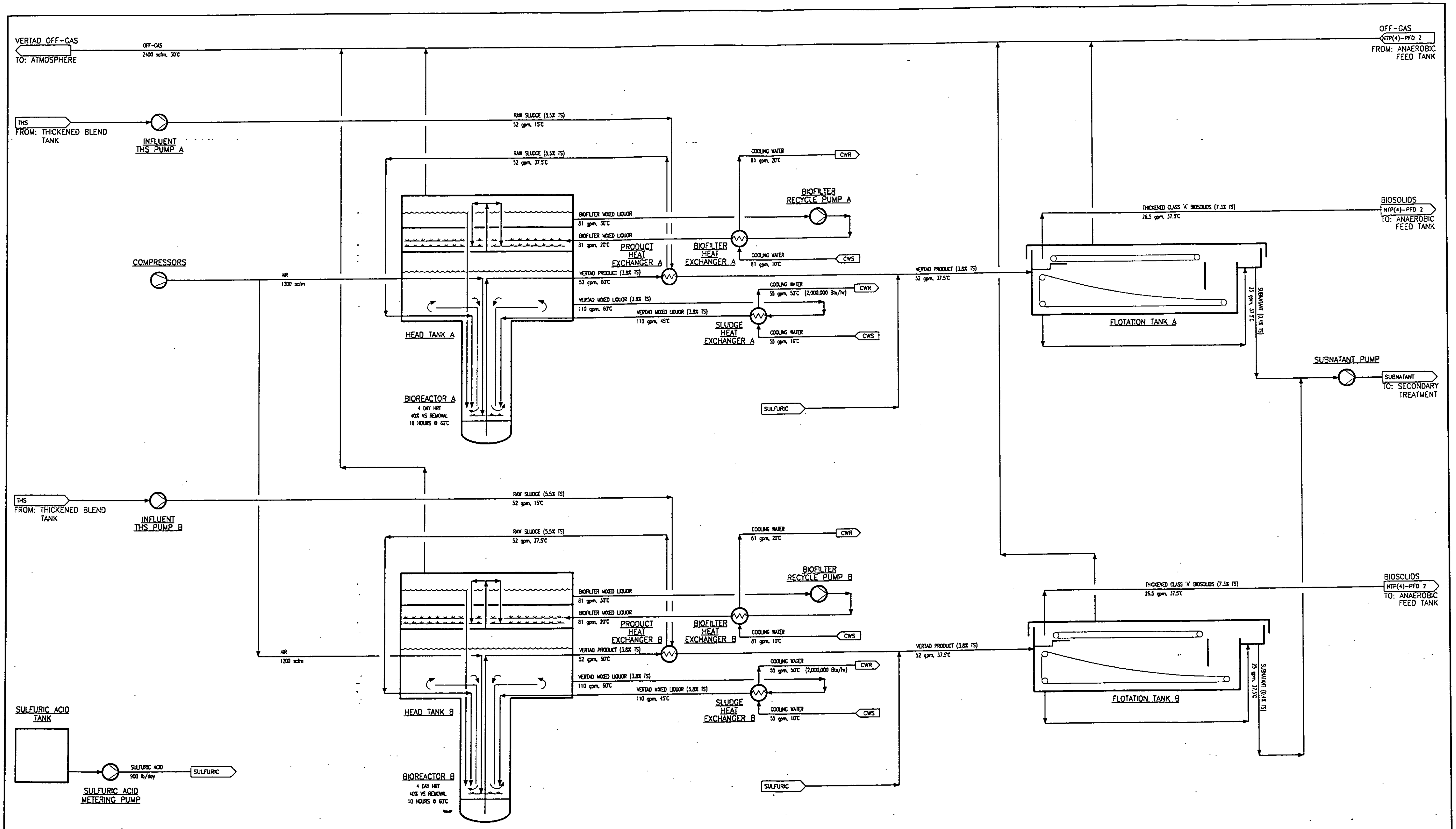
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
EAST SECTION RECLAMATION PLANT
PEAK WEEK-4 DAY HRT VERTAD ONLY

PROJECT NO	DRAWING NO	REV
	ESRP(V4)-PFD 3	A



Process Flow Diagram for Alternative NP1
Figure 3-7 1 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

- PUMP
- BLOWER
- HEAT EXCHANGER
- COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADDED REDUNDANCY	04JUL00	PR	JG		

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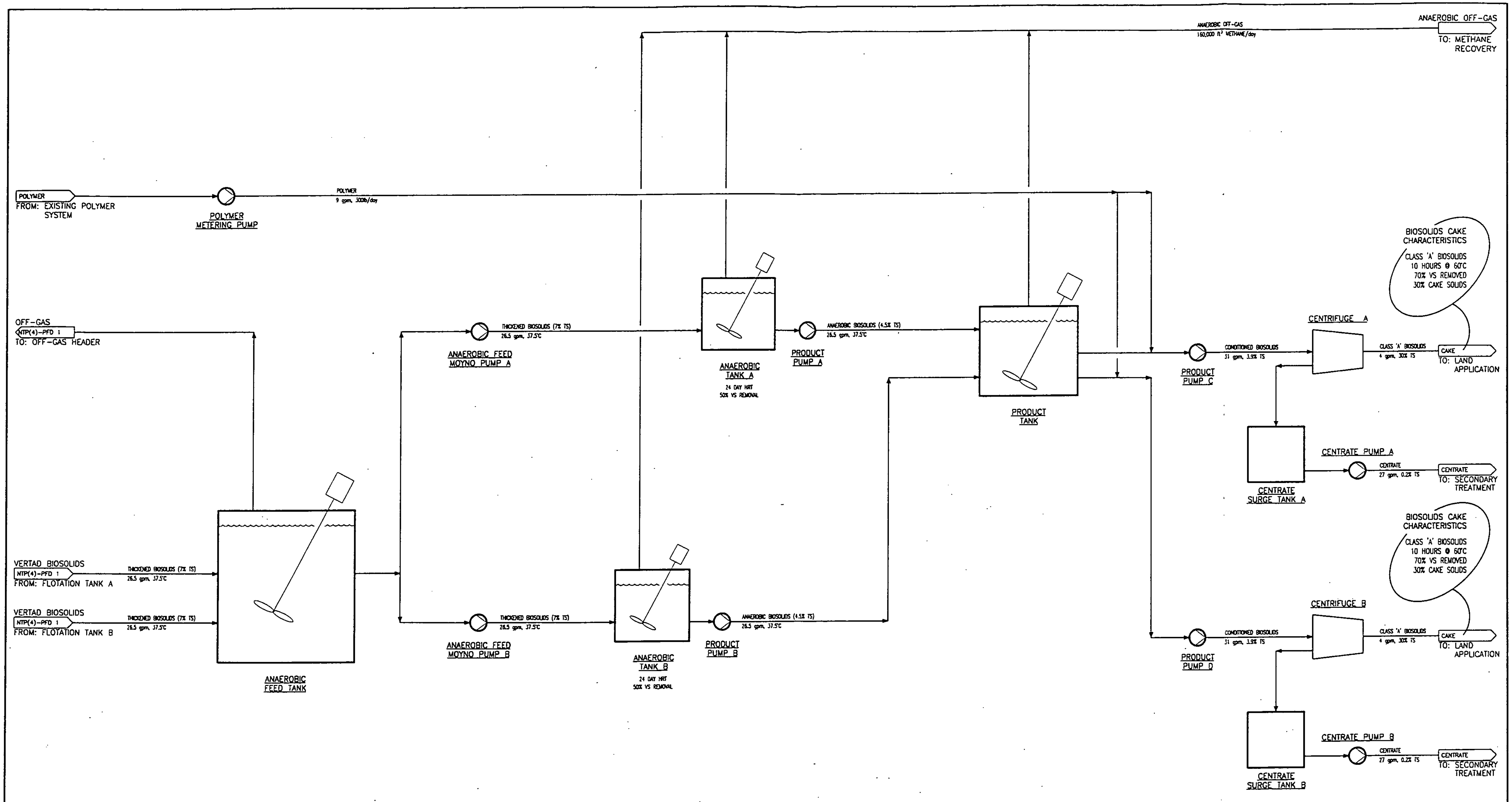
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN99	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
NORTH TREATMENT PLANT
4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	NTP(4)-PFD 1	B



Process Flow Diagram for Alternative NP1
Figure 3-7 2 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV.	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADDED REDUNDANCY	04JUL00	PR	JG		

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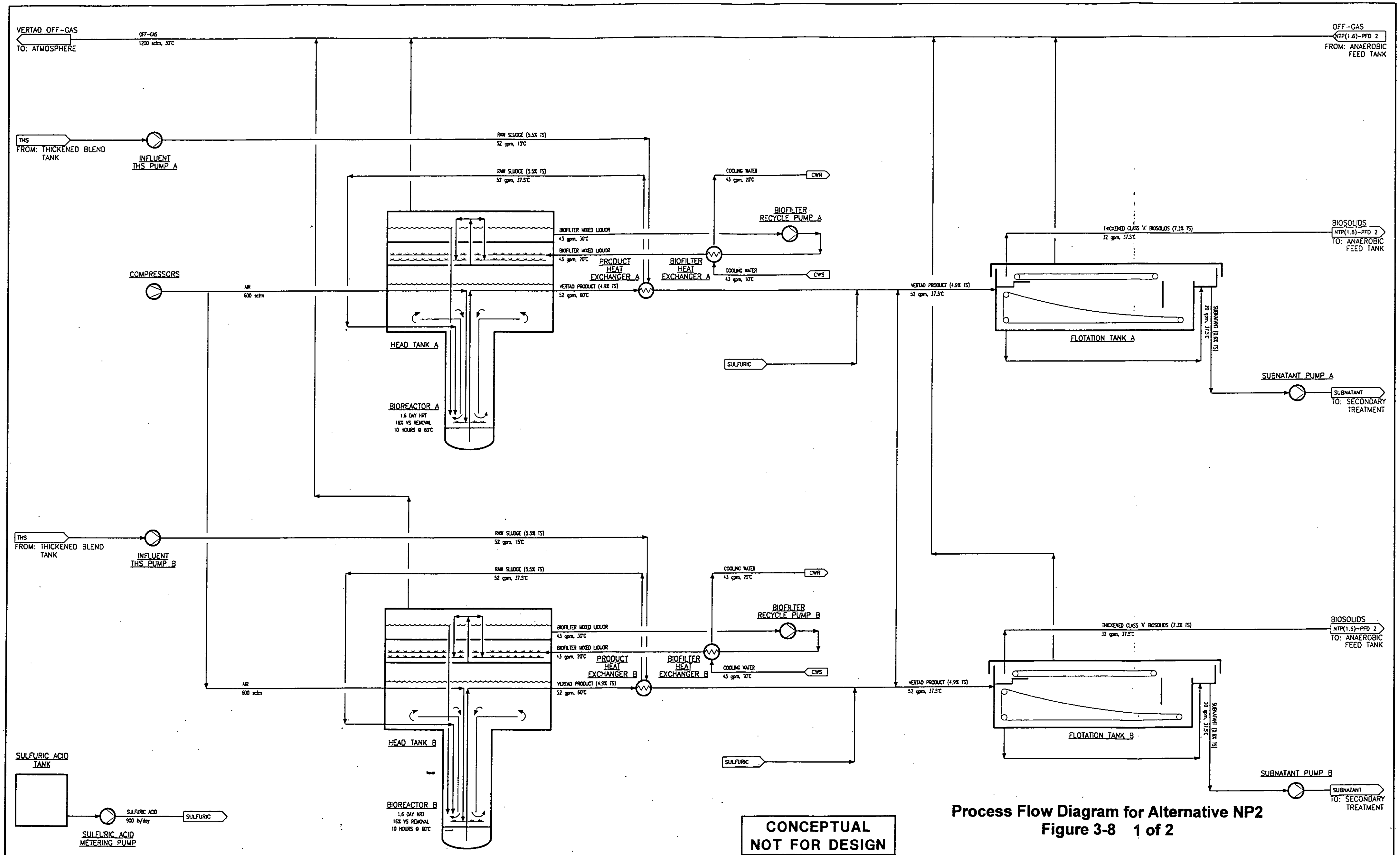
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
NORTH TREATMENT PLANT
4 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	NTP(4)-PFD 2	B



CONCEPTUAL NOT FOR DESIGN

Process Flow Diagram for Alternative NP2
Figure 3-8 1 of 2

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

- ○ PUMP
- BLOWER
- HEAT EXCHANGER
- COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADDED REDUNDANCY	04JUL00	PR	JG		

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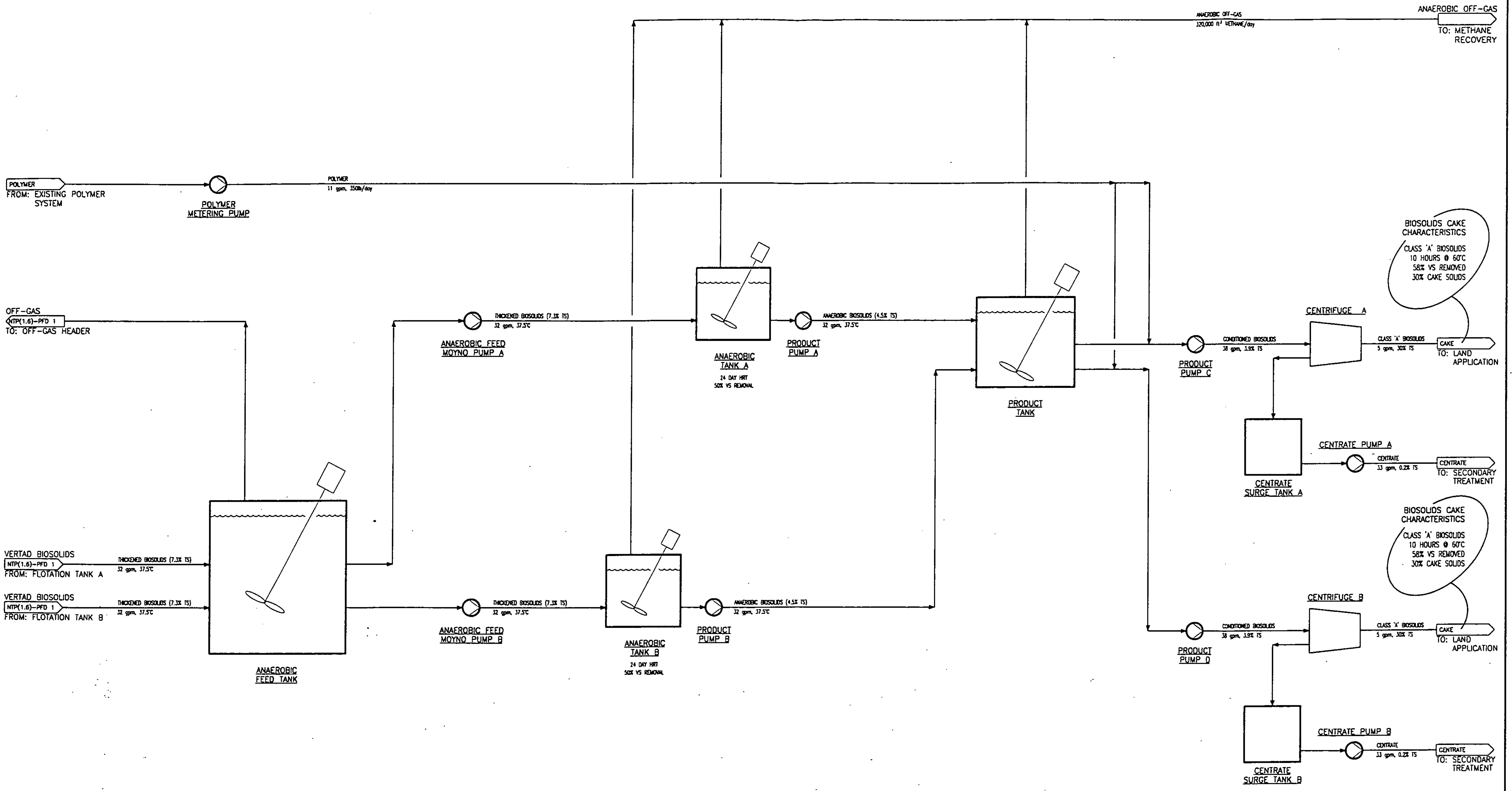
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	08JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
NORTH TREATMENT PLANT
1.6 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	NTP(1.6)-PFD 1	B



Process Flow Diagram for Alternative NP2
Figure 3-8 2 of 2

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

○ PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADDED REDUNDANCY	04JUL00	PR	JG		

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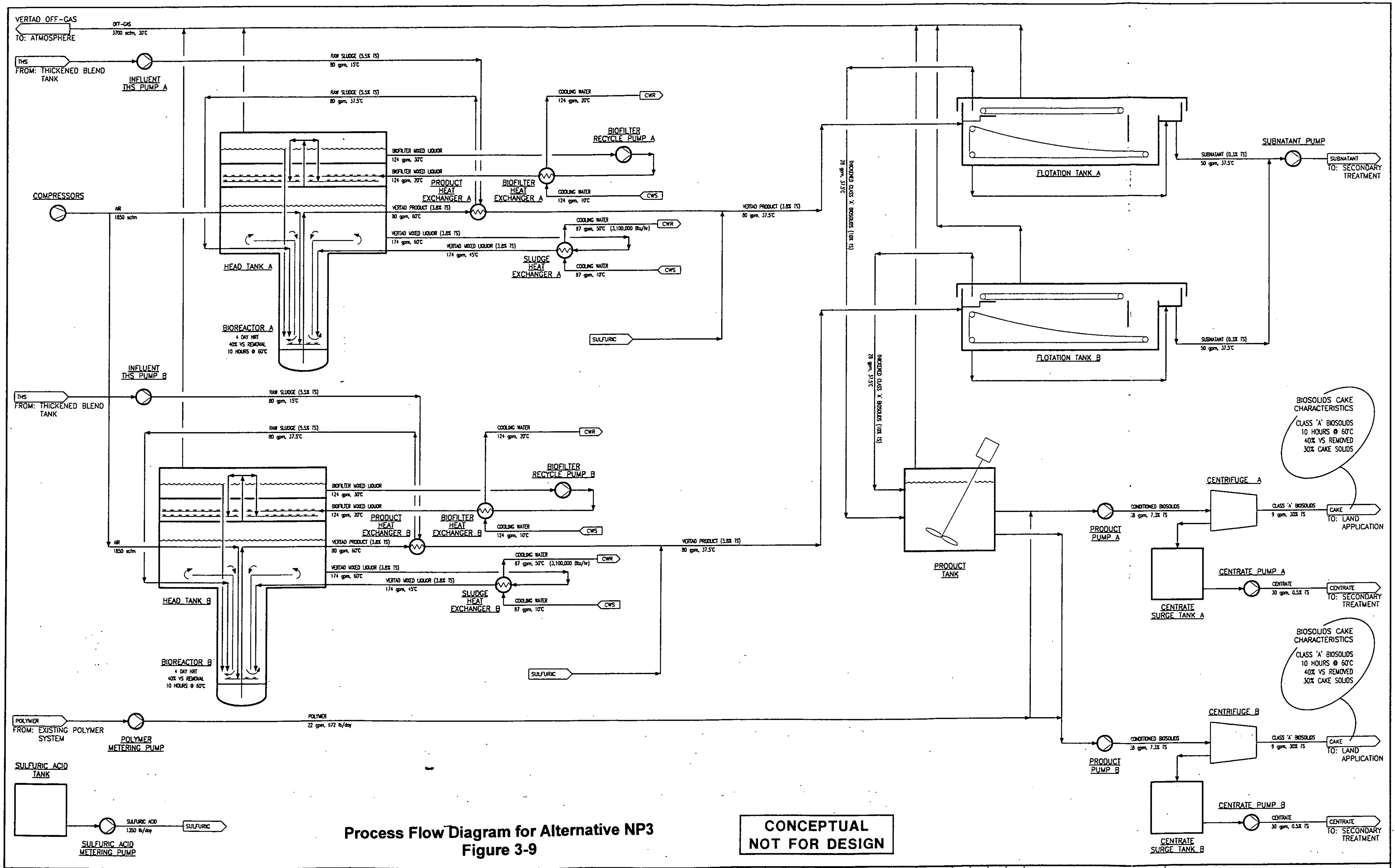
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	08JAN99	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
NORTH TREATMENT PLANT
1.6 DAY HRT VERTAD

PROJECT NO	DRAWING NO	REV
	NTP(1.6)-PFD 2	B



Process Flow Diagram for Alternative NP3
Figure 3-9

CONCEPTUAL
NOT FOR DESIGN

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

PUMP
BLOWER
HEAT EXCHANGER
COMPRESSOR

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	16MAY00	PR	JG		
B	ADDED REDUNDANCY	04JUL00	PR	JG		

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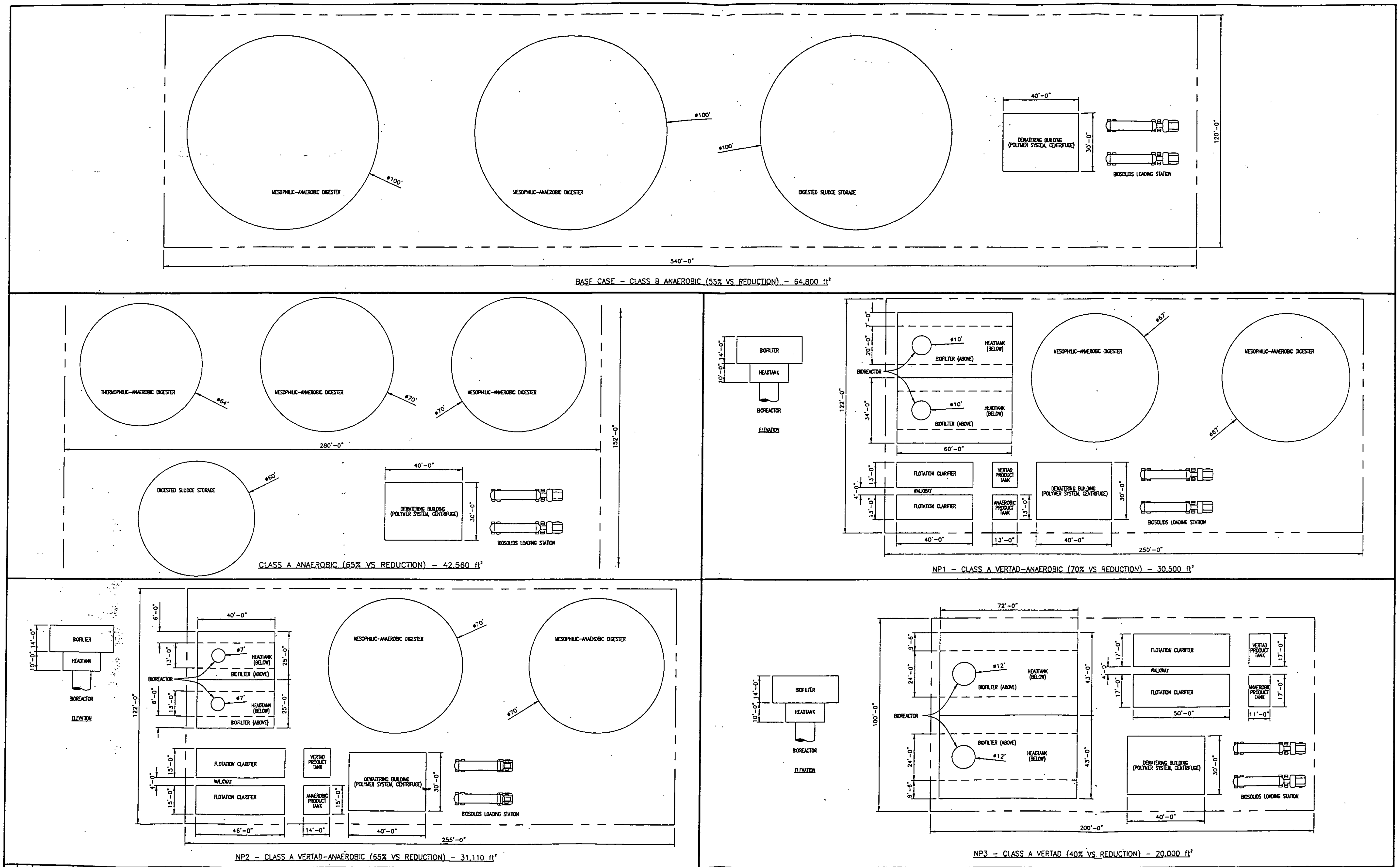
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	27APR00	PR
DES CHK		
DESIGN	06JAN00	JG

KING COUNTY

VERTAD SYSTEM CONCEPTUAL DESIGN
NORTH TREATMENT PLANT PEAK WEEK
4 DAY HRT VERTAD ONLY

PROJECT NO	DRAWING NO	REV
	NTP(V4)-PFD 1	B



BASE CASE - CLASS B ANAEROBIC (55% VS REDUCTION) - 64,800 ft²

CLASS A ANAEROBIC (65% VS REDUCTION) - 42,560 ft²

NP1 - CLASS A VERTAD-ANAEROBIC (70% VS REDUCTION) - 30,500 ft²

NP2 - CLASS A VERTAD-ANAEROBIC (65% VS REDUCTION) - 31,110 ft²

NP3 - CLASS A VERTAD (40% VS REDUCTION) - 20,000 ft²

North Treatment Plant Space Alternatives
Figure 3-10

REFERENCE DRAWINGS

NO	DWG NO	TITLE

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
A	PRELIMINARY	30AUG00	BB	JG		
01	FOR E & A REPORT	31AUG00	BB	JG		

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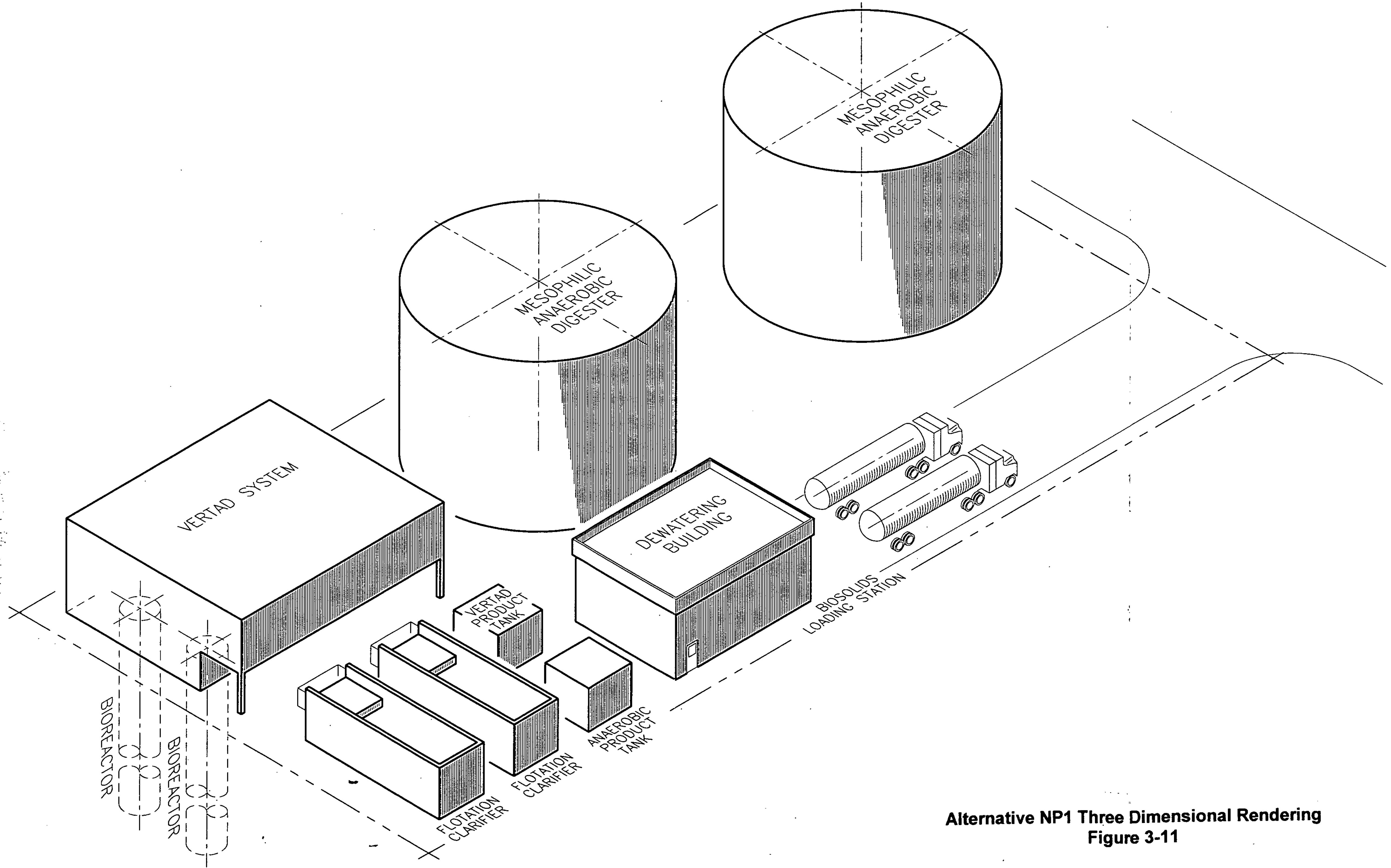
SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	30AUG00	BB
DES CHK		
DESIGN	30AUG00	JG

KING COUNTY

NORTH TREATMENT PLANT ALTERNATIVES

PROJECT NO	DRAWING NO	REV
B0312	NTP_SITE_LAYOUT	01



Alternative NP1 Three Dimensional Rendering
Figure 3-11

REFERENCE DRAWINGS

NO	DWG NO	TITLE

NOTES:

DRAWING REVISIONS

REV	DESCRIPTION	DATE	DRN	DES	CHK	P MGR
01	FOR E & A REPORT	05SEP00	PR	JG		

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Telephone (604)681-2030 Fax (604)683-9164

SCALE:	NTS	
PROJ MGR		
DWG CHK		
DRAWN	30AUG00	BB
DES CHK		
DESIGN	30AUG00	JG

KING COUNTY

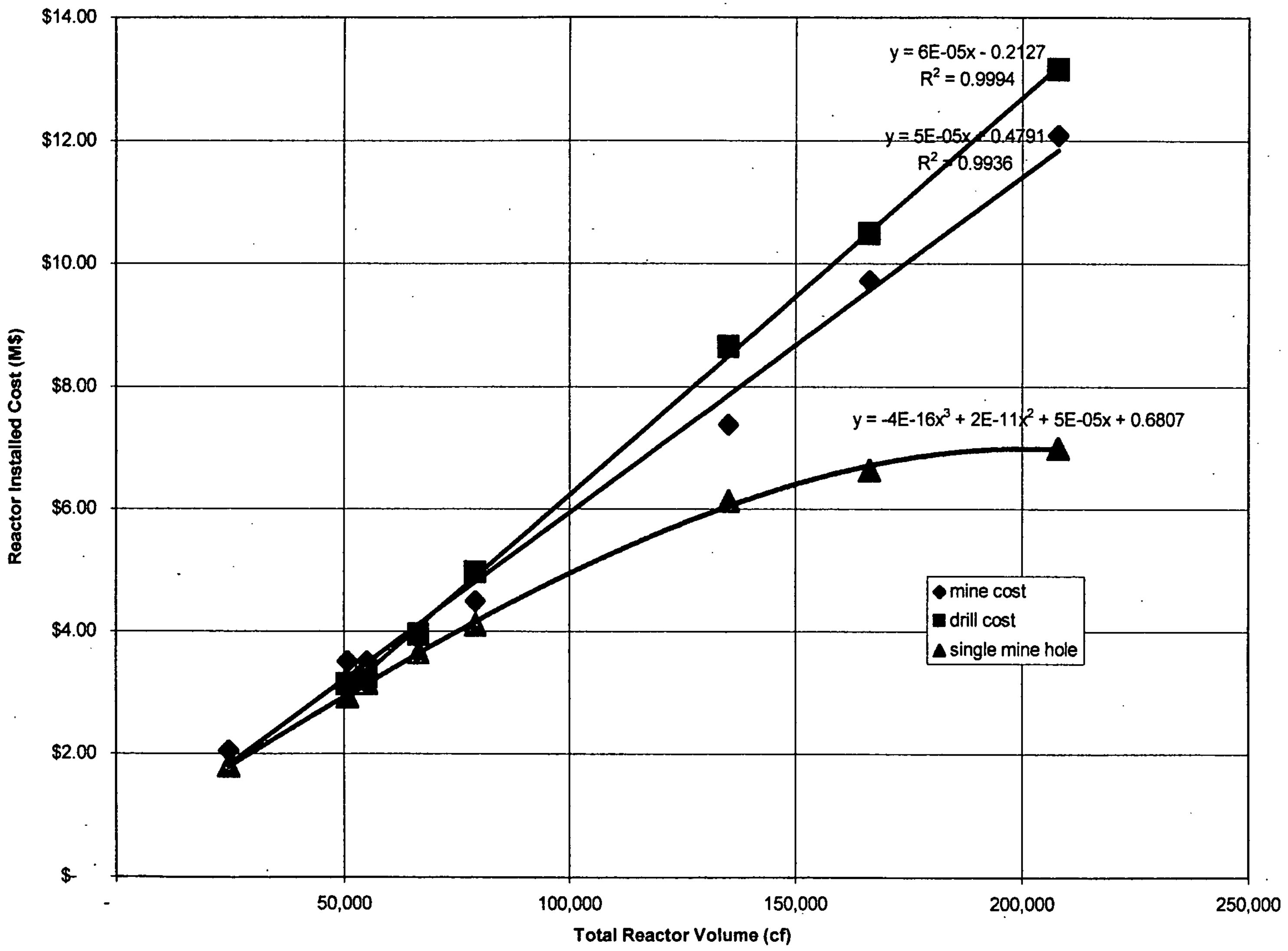
NORTH TREATMENT PLANT
ORTHO

PROJECT NO	DRAWING NO	REV
B0312	NP1 ORTHO	01

APPENDIX C

REACTOR CONSTRUCTION COST DATA

Drilling and Mining Cost Comparison



APPENDIX D

COST ESTIMATE SPREADSHEETS

Basic Cost Model by Brown and Caldwell

VERTAD™ Design Spreadsheets by NORAM

Cost Estimates and Cost Model Adaptation by E&A and King County

APPENDIX D (a)

North Treatment Plant Cost Estimates

North Treatment Plant VERTAD Alternative Analysis

COMMON TO ALL ALTERNATIVES

Planning Period	yr	2019
Influent Waste Water Flow	mgd	36
Peaking Factors		
Annual Average		1
Peak 3-week		1.22
Peak Week		1.52
Peak Day		1.66
Solids Unit Production Rate		
Primary Sludge	lbs-tss/MG	1310
Waste Activated Sludge	lbs-tss/MG	770
Thickened Sludge	lbs-tss/MG	1870
Solids Concentration		
Primary Sludge	percent solids	0.6%
Waste Activated Sludge	percent solids	0.5%
Thickened Sludge	percent solids	5.5%
Volatile Solids Ratio		
Primary Sludge	percent of tss	80%
Waste Activated Sludge	percent of tss	76%
Thickened Sludge	percent of tss	79%
Summer Sludge Temperature	F	71
Winter Sludge Temperature	F	57
Thermophilic Operating Temperature	F	135
Mesophilic Operating Temperature	F	95
Digester Cone Volume Percent Active	percent	75%
Digester Gas Production	(cf/lb vs dest)	15

Table 1.2
Projected Solids Process Flows and Loads
EDRP Digester 5 Predesign

Raw Solids	Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Primary Sludge				
Average Annual	654	942,446	47,160	37,728
Peak 3-Week	805	1,159,073	58,000	46,400
Peak Day	1,082	1,558,753	78,000	62,400
Waste Activated Sludge				
Average Annual	462	664,748	27,720	21,067
Peak 3-Week	563	810,993	33,818	25,702
Peak Day	766	1,103,482	46,015	34,972
Mixed Sludge				
Average Annual	1,116	1,607,194	74,880	58,795
Peak 3-Week	1,368	1,970,066	91,818	72,102
Peak Day	1,849	2,662,235	124,015	97,372
Thickening				

DAFTS		
Number of Units	no.	4
Diameter	ft	55
Number of Units	no.	2
Diameter	ft	65
Total Surface Area	sft	16,140
Solids Loading, All DAFT's in service		
Average Annual	lb/sf/day	4.6
Peak 3-Week	lb/sf/day	5.7
Peak Day	lb/sf/day	7.7
Solids Loading, one DAFT out of service		
Average Annual	lb/sf/day	5.8
Peak 3-Week	lb/sf/day	7.2
Peak Day	lb/sf/day	9.7

Thickened Solids	Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Thickened Sludge				
Average Annual	102	146,763	67,320	53,183
Peak 3-Week	124	179,050	82,130	64,883
Peak Day	169	243,626	111,751	88,283
			TSS DT/day	VSS DT/day
Average Annual			34	27
Peak 3-Week			41	32
Peak Week			52	40
Peak Day			56	44

Thickened Sludge Blending Tank		
Number of Units	no.	1
Diameter	ft	8
Sidewater Depth	ft	15
Tank Volume	gallons	5,640
Total Volume	gallons	5,640
Detention Time		
Average Annual	minutes	55.3
Peak 3-Week	minutes	45.4
Peak Day	minutes	33.3

North Treatment Plant VERTAD Alternative Analysis

ANAEROBIC DIGESTION AND CENTRIFUGE DEWATERING

Volatile Solids Reduction in Digesters	percent	55%
Dewatering Solids Capture	percent	92.5%
Dewatered Cake Solids	percent solids	23.40%

Anaerobic Mesophillic Digesters					
Existing					
Number of Units (existing)	no.	2			
Diameter	ft	100			
Sidewater Depth	ft	35.0			
Tank Cylindrical Volume	gallons	2,056,172			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75% of cone)	gallons	2,231,738			
Total Existing Volume	gallons	4,463,477			
Total Volume	gallons	4,463,477			
Detention Time					
All Digesters in Service					
Average Annual	days	30.4			
Peak 3-Week	days	24.9			
Peak Day	days	18.3			
One Digester Out of Service					
Average Annual	days	15.2			
Peak 3-Week	days	12.5			
Peak Day	days	9.2			
Volatile Solids Loading					
All Digesters in Service					
Average Annual	lbs-vs/cf/day	0.09			
Peak 3-Week	lbs-vs/cf/day	0.11			
Peak Day	lbs-vs/cf/day	0.15			
One Digester Out of Service					
Average Annual	lbs-vs/cf/day	0.11			
Peak 3-Week	lbs-vs/cf/day	0.14			
Peak Day	lbs-vs/cf/day	0.18			
Heat Demand			S/N surface ratio		
Digester Heat Loss to Control Bldg	BTUH	62,543	0.33		
Digester Heat Loss from Shell	BTUH	166,649	0.57	common	
Average Annual	BTUH	2,665,591	729,140	1,936,451	
Peak 3-Week	BTUH	3,091,610			
Peak Day	BTUH	3,943,648			
Digester Gas Production					
Average Annual	cf/day	438,758			
Peak 3-Week	cf/day	535,285			
Peak Day	cf/day	728,338			
Digested Solids		Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Digested Sludge					
Average Annual		102	146,763	38,069	23,932
Peak 3-Week		124	179,050	46,445	29,197
Peak Day		169	243,626	63,195	39,728
Digested Sludge Storage					
Blending Storage Tank					
Diameter	ft	100			
Sidewater Depth	ft	38.0			
Tank Cylindrical Volume	gallons	2,232,416			
Tank Cone Volume	gallons	Not Included			
Effective Tank Volume (includes 0% of cone)	gallons	2,232,416			

Total BST Volume	gallons	2,232,416		
Maximum Storage Capacity				
Average Annual	days	15.2		
Peak 3-Week	days	12.5		
Peak Day	days	9.2		
Dewatering		Flow gpm	TSS lbs/day	Concentration % solids
Loading				
Average Annual		102	38,069	3.11%
Max 3-Week		124	46,445	3.11%
Max Week	⇨	155	57,866	3.11%
Peak Day		169	63,195	3.11%
Centrifuge Dewatering				
Number of Units Needed (max week + 1 unit)	no.	2.0		
Nominal Size	model	CP3074		
Dewatering Capacity/Centrifuge	lbs/hr	2,500		
Operating Units at Annual Average Condition	no.	0.6		
Operating Units at Max 3-week Condition	no.	0.8		
Operating Units at Max Week Condition	no.	1.0		
Operating Units at Peak Day Condition	no.	1.1		
Dewatered Biosolids				
Dry Solids TSS (lbs)				
Average Annual	lbs/day	35,214		
Peak 3-Week	lbs/day	42,961		
Peak Day	lbs/day	58,456		
Dry Solids TSS (tons)				
Average Annual	dt/d	18		
Peak 3-Week	dt/d	21		
Peak Week	dt/d	27		
Peak Day	dt/d	29		
Dry Solids VSS (lbs)				
Average Annual	lbs/day	22,137		
Peak 3-Week	lbs/day	27,008		
Peak Day	lbs/day	36,748		
Wet Cake (lbs)				
Average Annual	wet lbs/day	150,488		
Peak 3-Week	wet lbs/day	183,596		
Peak Day	wet lbs/day	249,810		
Wet Cake (tons)				
Average Annual	wet ton/day	75		
Peak 3-Week	wet ton/day	92		
Peak Day	wet ton/day	125		

North Treatment Plant VERTAD Alternative Analysis
THERMO-MESO DIGESTION W/ HOLD TANKS, 2 CENTRIFUGES

Thermo-Meso Digestion Assumptions

Total Volatile Solids Reduction in Digesters	percent	65.00%
Assumed contribution to total digestion		
Themophilic Phase	fraction	0.6
Mesophilic Phase	fraction	0.4
Dewatering Solids Capture	percent	92.5%
Centrifuge Dewatered Cake Solids	percent solids	23.0%
Centridry Dried Cake Solids	percent solids	55.0%
Dryed Product Capacity	lbs/hr	0

Thermo-Meso Digestion Flows and Loads

Anaerobic Thermophillic Digester					
Existing					
Number of Units (existing)	no.	1			
Diameter	ft	64			
Sidewater Depth	ft	30.5			
Tank Cylindrical Volume	gallons	733,924			
Tank Cone Volume	gallons	234,088			
Effective cone fraction		75%			
Effective Tank Volume (includes 75% of cone)	gallons	909,490			
Total Existing Volume	gallons	909,490			
Total Volume	gallons	909,490			
Detention Time					
Average Annual	days	6.2			
Peak 3-Week	days	5.1			
Peak Day	days	3.7			
Volatile Solids Loading					
Average Annual	lbs-vs/cf/day	0.44			
Peak 3-Week	lbs-vs/cf/day	0.53			
Peak Day	lbs-vs/cf/day	0.73			
Heat Demand			S/N surface ratio		
Digester Heat Loss to Control Bldg	BTU/hr	33,505	0.33		
Digester Heat Loss from Shell of 1 Digester	BTU/hr	284,877	0.57		
Average Annual	BTU/hr	4,293,202			
Peak 3-Week	BTU/hr	5,167,663			
Peak Day	BTU/hr	6,916,584			
Thermophilic Gas Production					
Average Annual	cf/day	311,119			
Peak 3-Week	cf/day	379,566			
Peak Day	cf/day	516,458			
Thermophilic Digested Solids		Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Digested Sludge					
Average Annual		102	146,763	46,579	32,442
Peak 3-Week		124	179,050	56,826	39,579
Peak Day		169	243,626	77,321	53,853
Class A Hold Tanks					
Temp min temp for hold tank	F	135			
Temp min temp for hold tank	C	57.22			
Hold Time	hrs	11.7			
Size of 1 Hold Tank (at peak day)	gal	118,902	15,896 cf		

Rough Dimensions	diaxht	30	22.5 feet		
Estimated Heat Demand	BTU/hr	229,783			
Anaerobic Mesophillic Digesters					
Existing					
Number of Units (existing)	no.	2			
Diameter	ft	70			
Sidewater Depth	ft	41.0			
Tank Cylindrical Volume	gallons	1,180,243			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75% of cone)	gallons	1,355,809			
Total Existing Volume	gallons	2,711,618			
Total Volume	gallons	2,711,618			
Detention Time					
All Digesters in Service					
Average Annual	days	18.5			
Peak 3-Week	days	15.1			
Peak Day	days	11.1			
One Digester Out of Service, BST as a Digester					
Average Annual	days	18.6			
Peak 3-Week	days	15.2			
Peak Day	days	11.2			
Volatile Solids Loading					
All Digesters in Service					
Average Annual	lbs-vs/cf/day	0.09			
Peak 3-Week	lbs-vs/cf/day	0.11			
Peak Day	lbs-vs/cf/day	0.15			
One Digester Out of Service					
Average Annual	lbs-vs/cf/day	0.09			
Peak 3-Week	lbs-vs/cf/day	0.11			
Peak Day	lbs-vs/cf/day	0.15			
Cooling Demand					
Digester Heat Loss to Control Bldg	BTU/hr	142,144	Per Digester		
Digester Heat Loss from Shell	BTU/hr	179,445			
Average Annual	BTU/hr	-1,357,890	-452,630		
Peak 3-Week	BTU/hr	-1,806,332	-602,111		
Peak Day @ 135 F	BTU/hr	-2,703,214	-901,071		
Gas Production Meso Digestion					
Average Annual	cf/day	207,413			
Peak 3-Week	cf/day	253,044			
Peak Day	cf/day	344,305			
Digester Gas Production Total Digestion Process					
Average Annual	cf/day	518,532			
Peak 3-Week	cf/day	632,609			
Peak Day	cf/day	860,764			
Digested Solids		Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Digested Sludge					
Average Annual		102	146,763	32,751	18,614
Peak 3-Week		124	179,050	39,956	22,709
Peak Day		169	243,626	54,367	30,899
Digested Sludge Storage					
Blending Storage Tank					
Diameter	ft	60			
Sidewater Depth	ft	35.00			
Tank Cylindrical Volume	gallons	740,222			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75% of cone)	gallons	915,788			
Total BST Volume	gallons	915,788			
Maximum Storage Capacity (no cone volume considered)					
Average Annual	days	5.0			
Peak 3-Week	days	4.1			
Peak Day	days	3.0			

Dewatering		Flow gpm	TSS lbs/day	Concentration % solids
Loading				
Average Annual		102	32,751	2.68%
Max 3-Week		124	39,956	2.68%
Max Week		155	49,782	2.68%
Peak Day		169	54,367	2.68%
Centridry Dewatering/Drying				
Number of Units	no.	0.0		
Nominal Size	model	CD3074		
Dewatering Capacity	dry lbs/hr	0		
Heat Demand	BTU/hr	0		
Centrifuge Dewatering				
Number of Units Needed (Max Week + 1 unit)	no.	2.0		
Nominal Size	model	CP3074		
Hydraulic Capacity/BFP	gpm	2,500		
Operating Units at Annual Average Condition	no.	0.5		
Operating Units at Max 3-week Condition	no.	0.7		
Operating Units at Max Week Condition	no.	0.8		
Operating Units at Peak Day Condition	no.	0.9		
Dewatered Biosolids				
Dry Solids TSS (lbs)				
Average Annual	lbs/day	30,295		
Peak 3-Week	lbs/day	36,960		
Peak Day	lbs/day	50,289		
Dry Solids TSS (tons)				
Average Annual		15		
Peak 3-Week		18		
Peak Week		23		
Peak Day		25		
Dry Solids VSS (lbs)				
Average Annual	lbs/day	17,218		
Peak 3-Week	lbs/day	21,006		
Peak Day	lbs/day	28,582		
Wet Cake from Centridry (lbs)				
Average Annual	wet lbs/day	0		
Peak 3-Week	wet lbs/day	0		
Peak Day	wet lbs/day	0		
Wet Cake from Centrifuges (lbs)				
Average Annual	wet lbs/day	131,717		
Peak 3-Week	wet lbs/day	160,694		
Peak Day	wet lbs/day	218,650		
Wet Cake (lbs)				
Average Annual	wet lbs/day	131,717		
Peak 3-Week	wet lbs/day	160,694		
Peak Day	wet lbs/day	218,650		
Wet Cake (tons)				
Average Annual	wet ton/day	66		
Peak 3-Week	wet ton/day	80		
Peak Day	wet ton/day	109		

Thermo-Meso Digestion Capacity Extension Analysis

Thermophilic Digesters			
Evaluation Criteria			
1 Peak Day Thermophilic HRT >,= 3.5 days			
Volume	gal	909,490	
Max allowable Peak Day flow	gpd	259,854	
Max allowable average sludge flow =		156,539	governs

Convert to Annual Average influent	mgd	38
Year ESRP to Reach this flow	YR	2024
Mesophilic Digesters		
2 Peak Day flow w/ all Digesters in service, DT >, = 10 days		
Total Volume	gal	2,711,618
Acceptable peak sludge flow =	gpd	271,162
therefore average flow =	gpd	163,350
3 Max 3-week flow w/ 1 Digester out of service & the BST as Meso Digester, DT >,= 10		
Total Volume	gal	2,723,533
Acceptable peak sludge flow =	gpd	272,353
therefore average flow =	gpd	223,240

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

Design Criteria

Influent Specifications (for year ??)	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH_3 mg/l
Thickened Solids											
Average Annual	104	149,129	55,000	68,388	44,000	54,710	80,756	100,413	24,735	30,764	681

Variables

Influent					
Percentage of Total Solids to VERTAD	100.0%	%			
Diluted THS Concentration	5.5%	%	55,000	mg/l	
THS Temperature	15	C	59	F	
Dilution Water Temperature	10	C	50	F	
Volatile Percentage of Total Solids	80.0%	%			
Reactor(s)					
HRT	4	days			
Temperature	60	C	140	F	
Oxygen Transfer Efficiency	50.0%	%			1
Oxygen Requirement	1.4	lbs O_2/lb VS Destroyed			
VS Destruction	40.0%	%	1		
COD Destruction	50.0%	%			
FOG Destruction	90.0%	%			
Org-N Destruction	45.0%	%			
Heat Generation	9000	Btu/lb VS Destroyed			
Biofilter Loading	11.0	m^3/hr-m^2			
Biofilter Temperature	30	C	86	F	
Biofilter Off-gas Temperature	30	C	86	F	
Product Constituents					
Ammonia (NH_3)	1000	mg/l	1236	lb/day	
Ammonium Bicarbonate (NH_4HCO_3)	4,647	mg/l			
% Bicarbonate Release to Float to this %tage	25.0%	%			
Product Flow per VerTad	50	gpm			
VerTad Internal Solids Concentration	3.8%	%	38,411	mg/l	
Flotation Thickener					
Float Solids Concentration	7.0%	%	69,545	mg/l	
Capture Efficiency	95.0%	%			
Surface Solids Loading	1.8	lb/ft^2/hr	43.2	lb/ft^2/d	
93% Sulfuric Acid Addition	721	mg/l	0.0004	gal H_2SO_4/gal Product	
Polymer Split to Flotation Thickeners	0.0%	%			
Anaerobic Digester					
HRT	24	days			
Temperature	35	C	95	F	
VS Destruction	50.0%	%			
COD Destruction	50.0%	%			
FOG Destruction	50.0%	%			
Org-N Destruction	50.0%	%			
Internal TS concentration (Product)	4.5%	%	45,000	mg/l	
Gas Production	0.54	L CH_4/g COD_{rem}	(Jenny Yoo)		1
Heat of Combustion of Methane	22,773	Btu/lb CH4			
Specific Volume of Methane	24.2	cf/lb CH4			
Energy Constant	2546	Btu/hr/hp			
Digester Volume	914,257	gallons			
Centrifuge					
Cake Solids Concentration	30.0%	%	300,000	mg/l	
Capture Efficiency	95.0%	%			
Polymer Addition	20	lb/dry ton			
Polymer Concentration	0.26%	%	2,560	mg/l	
Geology					
Ground Temperature	10	C	50	F	
Overall Transfer Coeff. (Shaft to Dirt)	0.34	Btu/hr-°F-ft^2			
Overall Transfer Coeff. (Head Tank to Air)	N/A	Btu/hr-°F-ft^3			
Ambient Air					
Air Temperature	15	C	59	F	

Equipment Sizing (Average Annual)

Reactor(s)				
THS Flowrate	104	gpm	149,129	gpd
Dilution Water	0	gpm	0	gpd
Total Liquid Flow to Reactor(s)	104	gpm	149,129	gpd
Active Volume Required	79,743	ft^3		
Shaft				
Depth	350	ft		

Diameter	120 in	10.0 ft	
Volume	27,714 ft³		
Time Requirement at Reactor Temperature	287 min	4.8 hr	
Soak Zone Volume Required	1,986 ft³		
Soak Zone Depth	25.1 ft		
Soak Zone Safety Factor	1.1		
Actual Soak Zone Depth for Design	27.6 ft		
Actual Soak Zone Volume	2185 ft³		
Time in Soak Zone	316 min	5.3 hr	
Head Tank			
Sidewater Depth	10.0 ft		
Width	20.1 ft		
Length	60.2 ft		
Head Tank Surface Area	1,210 ft²		
Active Volume	12,148 ft³		
Total Active Volume per Reactor	39,862 ft³		
Number of Reactors Required	2.00		

Total VerTad Digester Volume:
596,514 gallons

Goal Seek
Set cell to an even number. Solve for shaft diameter cell above (repeat for desired shaft diameter)

Compressor(s)

Percentage of Energy Recoverable from Compressor	20%		
TS Loading on the Shaft(s)	68,388 lb/day		1294 kg/hr
Total VS Destroyed	21,881 lb/day		414 kg/hr
Total Oxygen Requirement	30,638 lb/day	368,918 ft³/day	
OTE	50%		
Total Aeration Requirement	3,513,503 ft³/day	264,215 lb/day	
Total Aeration Rate	2,440 scfm	678 hp	505 kW
Total Aeration Rate per Shaft	1220 scfm		
Compressed Air Temperature	32 C	90 F	

Available energy with 20% recovery:
345,074 Btu/hr

1.220859 kWhr/kg VS destroyed
0.390675 kWhr/kg TS in

Voidage Check

Voidage	0.63753 ft³ per scfm of aeration (at 14 scfm)	
Total Voidage in the Bioreactor(s) plus Head Tank(s)	1,562 ft³	2.0% of the active volume
Shaft Cross-sectional Area	79.2 ft²	
Riser Cross-sectional Area	59.4 ft²	
Downcomer Cross-sectional Area	19.8 ft²	
Riser Liquid Velocity	2.5 ft/s	
Bubble Rise Velocity	1.0 ft/s	
Bulk Riser Velocity	3.5 ft/s	
Riser Flowrate	12,471 ft³/min	
Aeration at Top of Bioreactor	1,040 scfm	
Voidage at the Top of a Bioreactor	8.3% %	(Stay below 14%)
Voidage at the Top of a Bioreactor	15.4 scfm/ft²	(Stay below 40scfm/ft²)

Biofilter(s)

Total Aeration Rate to Shaft(s)	2,440 scfm	
Biofilter Loading Rate	11.0 m³/hr-m²	0.601 ft/min
Total Biofilter Surface Area Required	4,056 ft²	
Biofilter Surface Area per Shaft	2,028 ft²	
Length of Biofilter	60.2 ft	
Width of Biofilter	33.7 ft	
Depth under Media	1 ft	
Media Depth	9 ft	
Standpipe Depth over Media	3 ft	
Active Volume per Biofilter (w/media)	26,361 ft³	
Biofilter Porosity	40% %	
Active Liquid Volume per Biofilter	10,544 ft³	
Total Volume of Biofilter(s)	52,734 ft³	
Total Condensation in the Biofilter(s)	2.73 gpm	
Condensation per Biofilter	1.36 gpm	

SAFT(s)

Total Off-Gas Flowrate	0.59 gpm	847 gpd	
Total Product Flowrate	103 gpm	148,281 gpd	
Product Concentration	3.8% %	38,411 mg/l	
Total Sulfuric Flowrate	0.04 gpm	58 gpd	0.44 ton/day
Total Liquid Flow to SAFT(s)	103 gpm	148,340 gpd	
TS Loading to the SAFT(s)	46,504 lb/day	1,938 lb/hr	
Surface Area Required	1,076 ft²		
Number of SAFTs Required	2		
Surface Area Required per SAFT	538 ft²		
Width	13.4 ft	(Stay below 20ft)	
Length	40.2 ft		
Sidewater Depth	12.0 ft		
Free-board	1.0 ft		
Active Volume per SAFT	6,457 ft³		
Total Volume of SAFT(s)	13,994 ft³		
HRT of SAFT(s)	17 hrs		

Ratio of Volatile to Total Solids in VerTad

5.5%	68,388 lb/day	TSin
1.1%	13677.581 lb/day	FSin
4.4%	54710.323 lb/day	VSin
	0.8	VS/TSin
2.6%	32826.194 lb/day	VSout
3.7%	46503.775 lb/day	TSout
1.1%	13677.531 lb/day	FSout
	0.7058824	VS/TSout

Product Storage Tank (SAFT Float Solids)

Total Liquid Flow to SAFT(s)	103 gpm	148,340 gpd
TS Loading to the SAFT(s)	46,504 lb/day	1,938 lb/hr
TS Subnatant Return from SAFT(s)	2,325 lb/day	97 lb/hr

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Thickened Solids (TS) from SAFT(s)	44,179 lb/day	1,841 lb/hr
Thickened Solids from SAFT(s)	53 gpm	76,188 gpd
Volatile Solids from SAFT(s)	31,185 lb/day	1,299 lb/hr
Total Subnatant Return from SAFT(s)	50 gpm	72,152 gpd
Subnatant Return Solids Concentration	3,865 mg/l	0.39% %
Underflow Subnatant Return from SAFT(s)	**15% %**	(Set by standpipe height in the subnatant trough)
HRT of Storage Tank(s)	4 hrs	
Total Active Storage Tank Volume Required	1,697 ft³	
Tank Height	**11.0 ft**	
Free-board	**1.0 ft**	
Maximum Sidewater Depth	10.0 ft	
Total Surface Area of Storage Tank(s)	170 ft²	
Number of Storage Tanks Required	**1**	
Width of Storage Tank(s)	13 ft	
Length of Storage Tank(s)	12.7 ft	

Anaerobic Digester(s)

Total Liquid Flow to Digester(s)	53 gpm	76,188 gpd
Total TS Loading to the Digester(s)	14,179 lb/day	1,841 lb/hr
Total VS Loading to the Digester(s)	31,185 lb/day	1,299 lb/hr
Number of Digesters Required	**2**	
Liquid Flow per Digester	26 gpm	38,094 gpd
TS Loading per Digester	22,089 lb/day	920 lb/hr
VS Loading per Digester	15,592 lb/day	650 lb/hr
Influent Solids Concentration	7.0% %	69,545 mg/l
Desired Internal Solids Concentration	4.5% %	45,000 mg/l
Desired Digester HRT	24.0 days	
Actual Digester HRT	24.0 days	
VS out of the Anaerobic Digester	7,796 lb/day	325 lb/hr
TS out of the Anaerobic Digester	14,293 lb/day	596 lb/hr
Liquid Flow per Digester	26 gpm	38,094 gpd
Product Solids Concentration	4.5% %	45,000 mg/l
Total Digester Volume Requirement	1,828,514 gallons	
Active Volume per Digester	914,257 gallons	
Total Methane Production	6,154,333 L/day CH4	
Total Methane Production	217,339 cf/day CH4	
Total Combined VS Destruction	38,297 lbVS dest/day	
Total Methane Production	5.7 cf CH4/lb VS dest	
Heat Available from Methane	204,519,143 Btu/day	
Heat Available from Methane	8,521,631 Btu/hr	
Methane per Anaerobic Digester	3,077,167 L/day CH4	
Overall Combined VS Destruction	70.0%	
TS Loading from the Anaerobic Digester	28,586 lb/day	1,191 lb/hr

Goal Seek
Set cell equal to value in cell E196,
and adjust cell D61

Goal Seek
Set cell equal to value in cell D56,
and adjust cell D42

c.f. methane from straight anaerobic:
13,676,297 L/day CH4
482,977 cf/day CH4
32,826 lbVS dest/day
14.7 cf CH4/lb VS dest
454,486,985 Btu/day
18,936,958 Btu/hr

Anaerobic Product Storage Tank

Total Liquid Flow to Tank	53 gpm	76,188 gpd
Anaerobic Product Solids Concentration	4.5% %	
HRT of Storage Tank(s)	4 hrs	
Total Active Storage Tank Volume Required	1,697 ft³	
Tank Height	**11.0 ft**	
Free-board	**1.0 ft**	
Maximum Sidewater Depth	10.0 ft	
Total Surface Area of Storage Tank(s)	170 ft²	
Number of Storage Tanks Required	**1**	
Width of Storage Tank(s)	13 ft	
Length of Storage Tank(s)	12.7 ft	

Centrifuge(s)

Total Solids from Anaerobic Digester(s)	28,586 lb/day	1,191 lb/hr
Total Solids from Anaerobic Digester(s)	53 gpm	76,188 gpd
Polymer Addition	20.0 lb/dry ton	
Made-down Polymer Flowrate	9 gpm	13,392 gpd
Polymer Flowrate	286 lb/day	12 lb/hr
Total Flow to Centrifuge(s)	62 gpm	89,581 gpd
Concentration of Flow to Centrifuge(s)	38,655 mg/l	3.9% %
Total Mass Loading on Centrifuge(s)	28,372 lb/day	1,203 lb/hr
Total Centrate Solids	1,444 lb/day	60 lb/hr
Total Cake Solids	27,423 lb/day	14 ton/day
Total Wet Cake Solids	91,428 Wet lb/day	46 Wet ton/day
Total Cake Flow to Trucks	7.6 gpm	10,965 gpd
Cake Solids Concentration	300,000 mg/l	30% %
Total Centrate Flow to Plant	55 gpm	78,615 gpd
Centrate Solids Concentration	2,292 mg/l	0.22% %

OTHER

Centrate Surge Tank Size

Retention Time Required	10 min	
Centrate Flowrate	55 gpm	78,615 gpd
Total Volume Required	73 ft³	546 gallons
Number of Tanks	2	

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Volume Required per Tank	36 ft³	273 gallons
Tank Height	8 ft	
Tank Diameter	2.4 ft	
Sulfuric Acid Tank Size		
# of Days of Sulfuric Acid Supply Required	7 days	
Total Sulfuric Flowrate	0.04 gpm	58 gpd
Total Volume Required	55 ft³	408 gallons
Tank Height	10 ft	
Tank Diameter	2.6 ft	
Influent Preheat Heat Exchanger		
Heat Recovery per Shaft	1,044,583 Btu/hr	
Heat Transfer Coefficient	75 Btu/hr-°F-ft²	
Sludge to Sludge Temperature Approach	0 C	
Hot Product Supply	60.0 C	140 F
Cool Influent Sludge	15.0 C	59 F
Tempered Product Return	37.5 C	99.5 F
Preheated Sludge to Shaft	37.5 C	99.5 F
Log Mean Differential Temperature	22.5 C	41 F
# of Heat Exchangers	2	
Surface Area for each Heat Exchanger	344 ft²	
Flowrate of product sludge through exchanger	51 gpm	
Shaft Internal Recycle Heat Exchanger		
Cooling Requirement per Shaft	1,984,126 Btu/hr	
Heat Transfer Coefficient	75 Btu/hr-°F-ft²	
Sludge to Water Temperature Approach	5 C	
Sludge Supply	60.0 C	140 F
Cooling Water Supply	10.0 C	50 F
Sludge Return	45.0 C	113 F
Cooling Water Return	50.0 C	122 F
Log Mean Differential Temperature	20.0 C	36 F
# of Heat Exchangers	2	
Surface Area for each Heat Exchanger	736 ft²	
Flowrate of reactor sludge through exchanger	110 gpm	
Flowrate of water through exchanger	55 gpm	
Biofilter Heat Exchanger		
Cooling Requirement per Biofilter	733,657 Btu/hr	
Biofilter Temperature	30 C	
Heat Transfer Coefficient	75 Btu/hr-°F-ft²	
Water to Water Temperature Approach	0 C	
Biofilter Supply	30.0 C	86 F
Cooling Water Supply	10.0 C	50 F
Biofilter Return	20.0 C	68 F
Cooling Water Return	20.0 C	68 F
Differential Temperature	10.0 C	18 F
Surface Area for each Heat Exchanger	543 ft²	
# of Heat Exchangers	2	
Flowrate of biofilter liquor through exchanger	81 gpm	
Flowrate of water through exchanger	81 gpm	
Turnover time in the biofilter	969 min	16.2 hrs
Energy Recovery and Generation		
Total Heat Recovery from VerTad System	7,526,455 Btu/hr	
Heat Generation from VerTad Internal Recycle	3,969,161 Btu/hr	(As 122°F cooling water return)
Heat Removal from Biofilter	1,167,651 Btu/hr	
Heat used to Preheat Sludge to VerTad	2,089,644 Btu/hr	
Heat available from the Compressors	345,074 Btu/hr	(As 185°F cooling water return)

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

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Design Criteria

Influent Specifications **(for year ??)**	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH_3 mg/l
Thickened Solids											
Average Annual	104	149,129	**55,000**	68,388	44,000	54,710	**80,756**	100,413	**24,735**	30,764	**681**

Variables

Influent

Percentage of Total Solids to VERTAD	**100.0%**	%			
Diluted THS Concentration	**5.5%**	%	55,000	mg/l	
THS Temperature	15	C	59	F	
Dilution Water Temperature	10	C	50	F	
Volatile Percentage of Total Solids	**80.0%**	%			

Reactor(s)

HRT	1.6	days			
Temperature	**60**	C	140	F	
Oxygen Transfer Efficiency	**40.0%**	%			
Oxygen Requirement	**1.4**	lbs O_2/lb VS Destroyed			
VS Destruction	**16%**	%	1		
COD Destruction	**21%**	%			
FOG Destruction	**90%**	%			
Org-N Destruction	**16%**	%			
Heat Generation	9000	Btu/lb VS Destroyed			
Biofilter Loading	11.0	$m^3/hr\text{-}m^2$			
Biofilter Temperature	**30**	C	86	F	
Biofilter Off-gas Temperature	**30**	C	86	F	

Product Constituents

Ammonia (NH_3)	**1000**	mg/l	1240	lb/day
Ammonium Bicarbonate (NH_4 HCO_3)	4,647	mg/l		
% Bicarbonate Release to Float to this %tage	**25.0%**	%		
Product Flow per VerTad	51	gpm		
VerTad Internal Solids Concentration	4.9%	%	48,603	mg/l

Flotation Thickener

Float Solids Concentration	**7.3%**	%	73,232	mg/l
Capture Efficiency	**95.0%**	%		
Surface Solids Loading	**1.8**	$lb/ft^2/hr$	43.2	$lb/ft^2/d$
93% Sulfuric Acid Addition	721	mg/l	0.0004	gal H_2SO_4/gal Product
Polymer Split to Flotation Thickeners	0.0%	%		

Anaerobic Digester

HRT	**24**	days			
Temperature	**35**	C	95	F	
VS Destruction	**58.5%**	%			
COD Destruction	**58.5%**	%			
FOG Destruction	**58.5%**	%			
Org-N Destruction	**58.5%**	%			
Internal TS concentration (Product)	**4.5%**	%	45,000	mg/l	
Gas Production	**0.54**	L CH_4/g COD_{rem}		(Jenny Yoo)	1
Heat of Combustion of Methane	**22,773**	Btu/lb CH4			
Specific Volume of Methane	**24.2**	cf/lb CH4			
Energy Constant	2546	Btu/hr/hp			
Digester Volume	**1,104,881**	gallons			

Centrifuge

Cake Solids Concentration	**30.0%**	%	300,000	mg/l
Capture Efficiency	**95.0%**	%		
Polymer Addition	20	lb/dry ton		
Polymer Concentration	**0.26%**	%	2,560	mg/l

Geology

Ground Temperature	**10**	C	50	F
Overall Transfer Coeff. (Shaft to Dirt)	**0.34**	$Btu/hr\text{-}^{\circ}F\text{-}ft^2$		
Overall Transfer Coeff. (Head Tank to Air)	**N/A**	$Btu/hr\text{-}^{\circ}F\text{-}ft^3$		

Ambient Air

Air Temperature	15	C	59	F

Equipment Sizing (Average Annual)

Reactor(s)

THS Flowrate	104	gpm	149.129	gpd

Dilution Water	0 gpm	0 gpd	
Total Liquid Flow to Reactor(s)	104 gpm	149,129 gpd	
Active Volume Required	31,879 ft^3		
Shaft			
Depth	350 ft		
Diameter	80 in	6.7 ft	
Volume	12,332 ft^3		
Time Requirement at Reactor Temperature	287 min	4.8 hr	
Soak Zone Volume Required	1,987 ft^3		
Soak Zone Depth	56.4 ft		
Soak Zone Safety Factor	1.1		
Actual Soak Zone Depth for Design	62.0 ft		
Actual Soak Zone Volume	2185 ft^3		
Time in Soak Zone	316 min	5.3 hr	
Head Tank			
Sidewater Depth	6.7 ft		
Width	13.4 ft		
Length	40.2 ft		
Head Tank Surface Area	538 ft^2		
Active Volume	3,606 ft^3		
Total Active Volume per Reactor	15,938 ft^3		
Number of Reactors Required	2.00		
Compressor(s)			
Percentage of Energy Recoverable from Compressor	20%		
TS Loading on the Shaft(s)	68,388 lb/day		1294 kg/hr
Total VS Destroyed	8,749 lb/day		165 kg/hr
Total Oxygen Requirement	12,248 lb/day	147,482 ft^3/day	
OTE	40%		
Total Aeration Requirement	1,755,739 ft^3/day	132,032 lb/day	
Total Aeration Rate	1,219 scfm	339 hp	253 kW
Total Aeration Rate per Shaft	610 scfm		
Compressed Air Temperature	32 C	90 F	
Voidage Check			
Voidage	0.63753 ft^3 per scfm of aeration (at 14 scfm)		
Total Voidage in the Bioreactor(s) plus Head Tank(s)	857 ft^3	2.7% of the active volume	
Shaft Cross-sectional Area	35.2 ft^2		
Riser Cross-sectional Area	26.4 ft^2		
Downcomer Cross-sectional Area	8.8 ft^2		
Riser Liquid Velocity	2.5 ft/s		
Bubble Rise Velocity	1.0 ft/s		
Bulk Riser Velocity	3.5 ft/s		
Riser Flowrate	5,549 ft^3/min		
Aeration at Top of Bioreactor	550 scfm		
Voidage at the Top of a Bioreactor	9.9% %	(Stay below 14%)	
Voidage at the Top of a Bioreactor	17.3 scfm/ft^2	(Stay below 40scfm/ft^2)	
Biofilter(s)			
Total Aeration Rate to Shaft(s)	1,219 scfm		
Biofilter Loading Rate	11.0 m^3/hr-m^2	0.601 ft/min	
Total Biofilter Surface Area Required	2,027 ft^2		
Biofilter Surface Area per Shaft	1,013 ft^2		
Length of Biofilter	40.2 ft		
Width of Biofilter	25.2 ft		
Depth under Media	1 ft		
Media Depth	9 ft		
Standpipe Depth over Media	3 ft		
Active Volume per Biofilter (w/media)	13,175 ft^3		
Biofilter Porosity	40% %		
Active Liquid Volume per Biofilter	5,270 ft^3		
Total Volume of Biofilter(s)	26,352 ft^3		
Total Condensation in the Biofilter(s)	1.36 gpm		
Condensation per Biofilter	0.68 gpm		
SAFT(s)			
Total Off-Gas Flowrate	0.29 gpm	423 gpd	
Total Product Flowrate	103 gpm	148,705 gpd	
Product Concentration	4.9% %	48,603 mg/l	
Total Sulfuric Flowrate	0.04 gpm	58 gpd	0.45 ton/day
Total Liquid Flow to SAFT(s)	103 gpm	148,764 gpd	
TS Loading to the SAFT(s)	59,639 lb/day	2,485 lb/hr	
Surface Area Required	1,381 ft^2		
Number of SAFTs Required	2		
Surface Area Required per SAFT	690 ft^2		
Width	15.2 ft	(Stay below 20ft)	
Length	45.5 ft		
Sidewater Depth	12.0 ft		

NB: Could use a single 15ft diameter reactor

Total VerTad Digester Volume:
238,468 gallons

Goal Seek
Set cell to an even number, Solve for shaft diameter
cell above (repeat for desired shaft diameter)

Available energy with 20% recovery:
172,438 Btu/hr

1.526074 kWhr/kg VS destroyed
0.195225 kWhr/kg TS in

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Ratio of Volatile to Total Solids in VerTad

5.5%	68,388 lb/day	TSin
1.1%	13677.531 lb/day	FSin
4.4%	54710.323 lb/day	VSin
	0.8	VS/TSin

3.7%	45961.715 lb/day	VSout
4.8%	59639.295 lb/day	TSout
1.1%	13677.581 lb/day	FSout
	0.7706616	VS/TSout

Free-board	1.0	ft		
Active Volume per SAFT	8,283	ft^3		
Total Volume of SAFT(s)	17,947	ft^3		
HRT of SAFT(s)	22	hrs		

Product Storage Tank (SAFT Float Solids)

Total Liquid Flow to SAFT(s)	103	gpm	148,764	gpd
TS Loading to the SAFT(s)	59,639	lb/day	2,485	lb/hr
TS Subnatant Return from SAFT(s)	2,982	lb/day	124	lb/hr
Thickened Solids (TS) from SAFT(s)	56,657	lb/day	2,361	lb/hr
Thickened Solids from SAFT(s)	64	gpm	92,790	gpd
Volatile Solids from SAFT(s)	43,664	lb/day	1,819	lb/hr
Total Subnatant Return from SAFT(s)	39	gpm	55,974	gpd
Subnatant Return Solids Concentration	6,389	mg/l	0.64%	%
Underflow Subnatant Return from SAFT(s)	**15%**	%	(Set by standpipe height in the subnatant trough)	
HRT of Storage Tank(s)	4	hrs		
Total Active Storage Tank Volume Required	2,067	ft^3		
Tank Height	11.0	ft		
Free-board	1.0	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	207	ft^2		
Number of Storage Tanks Required	1			
Width of Storage Tank(s)	15	ft		
Length of Storage Tank(s)	13.6	ft		

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Anaerobic Digester(s)

Total Liquid Flow to Digester(s)	64	gpm	92,790	gpd
Total TS Loading to the Digester(s)	56,657	lb/day	2,361	lb/hr
Total VS Loading to the Digester(s)	43,664	lb/day	1,819	lb/hr
Number of Digesters Required	2			
Liquid Flow per Digester	32	gpm	46,395	gpd
TS Loading per Digester	28,329	lb/day	1,180	lb/hr
VS Loading per Digester	21,832	lb/day	910	lb/hr
Influent Solids Concentration	7.3%	%	73,232	mg/l
Desired Internal Solids Concentration	4.5%	%	45,000	mg/l
Desired Digester HRT	24.0	days		
Actual Digester HRT	23.8	days		
VS out of the Anaerobic Digester	9,060	lb/day	378	lb/hr
TS out of the Anaerobic Digester	15,557	lb/day	648	lb/hr
Liquid Flow per Digester	32	gpm	46,395	gpd
Product Solids Concentration	4.0%	%	40,216	mg/l
Total Digester Volume Requirement	2,226,951	gallons		
Active Volume per Digester	1,113,475	gallons		
Total Methane Production	11,378,228	L/day CH4		
Total Methane Production	401,821	cf/day CH4		
Total Combined VS Destruction	35,636	lbVS dest/day		
Total Methane Production	11.3	cf CH4/lb VS dest		
Heat Available from Methane	378,118,202	Btu/day		
Heat Available from Methane	15,754,925	Btu/hr		
Methane per Anaerobic Digester	5,689,114	L/day CH4		
Overall Combined VS Destruction	65.1%			
TS Loading from the Anaerobic Digester	31,114	lb/day	1,296	lb/hr

Goal Seek
Set cell equal to value in cell E196, and adjust cell D61

Goal Seek
Set cell equal to value in cell D56, and adjust cell D42

c.f. methane from straight anaerobic:

13,676,297	L/day CH4
482,977	cf/day CH4
32,826	lbVS dest/day
14.7	cf CH4/lb VS dest
454,486,985	Btu/day
18,936,958	Btu/hr

Anaerobic Product Storage Tank

Total Liquid Flow to Tank	64	gpm	92,790	gpd
Anaerobic Product Solids Concentration	4.0%	%		
HRT of Storage Tank(s)	4	hrs		
Total Active Storage Tank Volume Required	2,067	ft^3		
Tank Height	11.0	ft		
Free-board	1.0	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	207	ft^2		
Number of Storage Tanks Required	1			
Width of Storage Tank(s)	15	ft		
Length of Storage Tank(s)	13.6	ft		

Centrifuge(s)

Total Solids from Anaerobic Digester(s)	31,114	lb/day	1,296	lb/hr
Total Solids from Anaerobic Digester(s)	64	gpm	92,790	gpd
Polymer Addition	20.0	lb/dry ton		
Made-down Polymer Flowrate	10	gpm	14,577	gpd
Polymer Flowrate	311	lb/day	13	lb/hr
Total Flow to Centrifuge(s)	75	gpm	107,366	gpd
Concentration of Flow to Centrifuge(s)	35,104	mg/l	3.5%	%
Total Mass Loading on Centrifuge(s)	31,425	lb/day	1,309	lb/hr
Total Centrate Solids	1,571	lb/day	65	lb/hr

Total Cake Solids	29,854 lb/day	15 ton/day
Total Wet Cake Solids	99,513 Wet lb/day	50 Wet ton/day
Total Cake Flow to Trucks	8.3 gpm	11,935 gpd
Cake Solids Concentration	300,000 mg/l	30% %
Total Centrate Flow to Plant	66 gpm	95,431 gpd
Centrate Solids Concentration	1,975 mg/l	0.20% %

OTHER

Centrate Surge Tank Size

Retention Time Required	10 min	
Centrate Flowrate	66 gpm	95,431 gpd
Total Volume Required	89 ft^3	663 gallons
Number of Tanks	1	
Volume Required per Tank	89 ft^3	663 gallons
Tank Height	8 ft	
Tank Diameter	3.8 ft	

Sulfuric Acid Tank Size

# of Days of Sulfuric Acid Supply Required	7 days	
Total Sulfuric Flowrate	0.04 gpm	58 gpd
Total Volume Required	55 ft^3	409 gallons
Tank Height	10 ft	
Tank Diameter	2.6 ft	

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Influent Preheat Heat Exchanger

Heat Recovery per Shaft	1,047,714 Btu/hr	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Sludge to Sludge Temperature Approach	0 C	
Hot Product Supply	60.0 C	140 F
Cool Influent Sludge	15.0 C	59 F
Tempered Product Return	37.5 C	99.5 F
Preheated Sludge to Shaft	37.5 C	99.5 F
Log Mean Differential Temperature	22.5 C	41 F
# of Heat Exchangers	2	
Surface Area for each Heat Exchanger	345 ft^2	
Flowrate of product sludge through exchanger	52 gpm	

Shaft Internal Recycle Heat Exchanger

Cooling Requirement per Shaft	0 Btu/hr	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Sludge to Water Temperature Approach	5 C	
Sludge Supply	60.0 C	140 F
Cooling Water Supply	10.0 C	50 F
Sludge Return	45.0 C	113 F
Cooling Water Return	50.0 C	122 F
Log Mean Differential Temperature	20.0 C	36 F
# of Heat Exchangers	2	
Surface Area for each Heat Exchanger	0 ft^2	
Flowrate of reactor sludge through exchanger	0 gpm	
Flowrate of water through exchanger	0 gpm	

Biofilter Heat Exchanger

Cooling Requirement per Biofilter	366,668 Btu/hr	
Biofilter Temperature	30 C	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Water to Water Temperature Approach	0 C	
Biofilter Supply	30.0 C	86 F
Cooling Water Supply	10.0 C	50 F
Biofilter Return	20.0 C	68 F
Cooling Water Return	20.0 C	68 F
Differential Temperature	10.0 C	18 F
Surface Area for each Heat Exchanger	271 ft^2	
# of Heat Exchangers	2	
Flowrate of biofilter liquor through exchanger	41 gpm	
Flowrate of water through exchanger	41 gpm	
Turnover time in the biofilter	969 min	16.2 hrs

Energy Recovery and Generation

Total Heat Recovery from VerTad System	2,829,019 Btu/hr	
Heat Generation from VerTad Internal Recycle	0 Btu/hr	(As 122°F cooling water return)
Heat Removal from Biofilter	733,403 Btu/hr	
Heat used to Preheat Sludge to VerTad	2,095,616 Btu/hr	
Heat available from the Compressors	172,438 Btu/hr	

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

Peaking Factor = 1.52

Design Criteria

Influent Specifications (for year ??)	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH_3 mg/l
Thickened Solids											
Average Annual	104	149,129	55,000	68,388	44,000	54,710	80,756	100,413	24,735	30,764	681
Peak	157	226,675	55,000	103,950	44,000	83,160	80,756	152,628	24,735	46,761	681

Variables

Influent

Percentage of Total Solids to VERTAD	100.0%	%		
Diluted THS Concentration	5.5%	%	55,000	mg/l
THS Temperature	15	C	59	F
Dilution Water Temperature	10	C	50	F
Volatile Percentage of Total Solids	80.0%	%		

Reactor(s)

HRT	4	days		
Temperature	60	C	140	F
Oxygen Transfer Efficiency	50.0%	%		
Oxygen Requirement	1.4	lbs O_2/lb VS Destroyed		
VS Destruction	40.0%	%		
COD Destruction	50.0%	%	1	
FOG Destruction	90.0%	%		
Org-N Destruction	45.0%	%		
Heat Generation	9000	Btu/lb VS Destroyed		
Biofilter Loading	11.0	$m^3/hr\text{-}m^2$		
Biofilter Temperature	30	C	86	F
Biofilter Off-gas Temperature	30	C	86	F

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Product Constituents

Ammonia (NH_3)	1000	mg/l	1879	lb/day
Ammonium Bicarbonate (NH_4 HCO_3)	4,647	mg/l		
% Bicarbonate Release to Float to this %tage	25.0%	%		
Product Flow per VerTad	77	gpm		
VerTad Internal Solids Concentration	3.8%	%	38,411	mg/l

Flotation Thickener

Float Solids Concentration	10.0%	%	100,000	mg/l
Capture Efficiency	95.0%	%		
Surface Solids Loading	1.8	$lb/ft^2/hr$	43.2	$lb/ft^2/d$
93% Sulfuric Acid Addition	721	mg/l	0.0004	gal H_2SO_4/gal Product
Polymer Split to Flotation Thickeners	0.0%	%		

Anaerobic Digester

HRT	24	days			
Temperature	35	C	95	F	
VS Destruction	50.0%	%	1		
COD Destruction	50.0%	%			
FOG Destruction	50.0%	%			
Org-N Destruction	50.0%	%			
Internal TS concentration (Product)	4.5%	%	45,000	mg/l	
Gas Production	0.54	L CH_4/g COD_{rem}	(Jenny Yoo)		1
Heat of Combustion of Methane	22,773	Btu/lb CH4			
Specific Volume of Methane	24.2	cf/lb CH4			
Energy Constant	2546	Btu/hr/hp			
Digester Volume	2,000,000	gallons			

Centrifuge

Cake Solids Concentration	30.0%	%	300,000	mg/l
Capture Efficiency	95.0%	%		
Polymer Addition	20	lb/dry ton		
Polymer Concentration	0.26%	%	2,560	mg/l

Geology

Ground Temperature	10	C	50	F
Overall Transfer Coeff. (Shaft to Dirt)	0.34	Btu/hr-°F-ft^2		
Overall Transfer Coeff. (Head Tank to Air)	N/A	Btu/hr-°F-ft^3		

Ambient Air

Air Temperature	15	C	59	F

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Equipment Sizing (Average Annual)

Reactor(s)

THS Flowrate	157 gpm	226,675 gpd
Dilution Water	0 gpm	0 gpd
Total Liquid Flow to Reactor(s)	157 gpm	226,675 gpd
Active Volume Required	121,209 ft^3	
Shaft		
Depth	350 ft	
Diameter	144 in	12.0 ft
Volume	39,731 ft^3	
Time Requirement at Reactor Temperature	287 min	4.8 hr
Soak Zone Volume Required	3,019 ft^3	
Soak Zone Depth	26.6 ft	
Soak Zone Safety Factor	1.1	
Actual Soak Zone Depth for Design	29.3 ft	
Actual Soak Zone Volume	3321 ft^3	
Time in Soak Zone	316 min	5.3 hr
Head Tank		
Sidewater Depth	12.0 ft	
Width	24.0 ft	
Length	72.1 ft	
Head Tank Surface Area	1,734 ft^2	
Active Volume	20,852 ft^3	
Total Active Volume per Reactor	60,583 ft^3	
Number of Reactors Required	2.00	

Total VerTad Digester Volume:
906,701 gallons

Goal Seek
Set cell to an even number. Solve for shaft diameter cell above (repeat for desired shaft diameter)

Compressor(s)

Percentage of Energy Recoverable from Compressor	20%		
TS Loading on the Shaft(s)	103,950 lb/day		1966 kg/hr
Total VS Destroyed	33,264 lb/day		629 kg/hr
Total Oxygen Requirement	46,569 lb/day	560,755 ft^3/day	
OTE	50%		
Total Aeration Requirement	5,340,524 ft^3/day	401,607 lb/day	
Total Aeration Rate	3,709 scfm	1030 hp	768 kW
Total Aeration Rate per Shaft	1854 scfm		
Compressed Air Temperature	32 C	90 F	

Available energy with 20% recovery:
524,513 Btu/hr

1.220859 kWhr/kg VS destroyed
0.390675 kWhr/kg TS in

Voidage Check

Voidage	0.63753 ft^3 per scfm of aeration (at 14 scfm)	
Total Voidage in the Bioreactor(s) plus Head Tank(s)	2,209 ft^3	1.8% of the active volume
Shaft Cross-sectional Area	113.5 ft^2	
Riser Cross-sectional Area	85.1 ft^2	
Downcomer Cross-sectional Area	28.4 ft^2	
Riser Liquid Velocity	2.5 ft/s	
Bubble Rise Velocity	1.0 ft/s	
Bulk Riser Velocity	3.5 ft/s	
Riser Flowrate	17,879 ft^3/min	
Aeration at Top of Bioreactor	1,526 scfm	
Voidage at the Top of a Bioreactor	8.5% %	(Stay below 14%)
Voidage at the Top of a Bioreactor	16.3 scfm/ft^2	(Stay below 40scfm/ft^2)

Biofilter(s)

Total Aeration Rate to Shaft(s)	3,709 scfm	
Biofilter Loading Rate	11.0 m^3/hr-m^2	0.601 ft/min
Total Biofilter Surface Area Required	6,166 ft^2	
Biofilter Surface Area per Shaft	3,082 ft^2	
Length of Biofilter	72.1 ft	
Width of Biofilter	42.7 ft	
Depth under Media	1 ft	
Media Depth	9 ft	
Standpipe Depth over Media	3 ft	
Active Volume per Biofilter (w/media)	40,064 ft^3	
Biofilter Porosity	40% %	
Active Liquid Volume per Biofilter	16,026 ft^3	
Total Volume of Biofilter(s)	80,156 ft^3	
Total Condensation in the Biofilter(s)	4.14 gpm	
Condensation per Biofilter	2.07 gpm	

SAFT(s)

Total Off-Gas Flowrate	0.89 gpm	1,288 gpd	
Total Product Flowrate	157 gpm	225,388 gpd	
Product Concentration	3.8% %	38,411 mg/l	
Total Sulfuric Flowrate	0.06 gpm	89 gpd	0.68 ton/day
Total Liquid Flow to SAFT(s)	157 gpm	225,476 gpd	
TS Loading to the SAFT(s)	70,686 lb/day	2,945 lb/hr	
Surface Area Required	1,636 ft^2		

Number of SAFTs Required	2				
Surface Area Required per SAFT	818	ft^2			
Width	16.5	ft	(Stay below 20ft)		
Length	49.5	ft			
Sidewater Depth	**12.0**	ft			
Free-board	**1.0**	ft			
Active Volume per SAFT	9,814	ft^3			
Total Volume of SAFT(s)	21,271	ft^3			
HRT of SAFT(s)	17	hrs			

Ratio of Volatile to Total Solids in VerTad

5.5%	103,950	lb/day	TSin
1.1%	20789.923	lb/day	FSin
4.4%	83159.691	lb/day	VSin
	0.8		VS/TSin
2.6%	49895.815	lb/day	VSout
3.7%	70685.738	lb/day	TSout
1.1%	20789.923	lb/day	FSout
	0.7058824		VS/TSout

Product Storage Tank (SAFT Float Solids)

Total Liquid Flow to SAFT(s)	157	gpm	225,476	gpd	
TS Loading to the SAFT(s)	70,686	lb/day	2,945	lb/hr	
TS Subnatant Return from SAFT(s)	3,534	lb/day	147	lb/hr	
Thickened Solids (TS) from SAFT(s)	67,151	lb/day	2,798	lb/hr	
Thickened Solids from SAFT(s)	56	gpm	80,538	gpd	
Volatile Solids from SAFT(s)	47,401	lb/day	1,975	lb/hr	
Total Subnatant Return from SAFT(s)	101	gpm	144,939	gpd	
Subnatant Return Solids Concentration	2,925	mg/l	0.29%	%	
Underflow Subnatant Return from SAFT(s)	**15%**	%			(Set by standpipe height in the subnatant trough)
HRT of Storage Tank(s)	**4**	hrs			
Total Active Storage Tank Volume Required	1,794	ft^3			
Tank Height	**11.0**	ft			
Free-board	**1.0**	ft			
Maximum Sidewater Depth	10.0	ft			
Total Surface Area of Storage Tank(s)	179	ft^2			
Number of Storage Tanks Required	**1**				
Width of Storage Tank(s)	17	ft			
Length of Storage Tank(s)	10.9	ft			

Centrifuge(s)

Total Solids from VerTad	67,151	lb/day	2,798	lb/hr
Total Solids from VerTad	56	gpm	80,538	gpd
Polymer Addition	20.0	lb/dry ton		
Made-down Polymer Flowrate	22	gpm	31,460	gpd
Polymer Flowrate	672	lb/day	28	lb/hr
Total Flow to Centrifuge(s)	78	gpm	111,998	gpd
Concentration of Flow to Centrifuge(s)	72,629	mg/l	7.3%	%
Total Mass Loading on Centrifuge(s)	67,823	lb/day	2,826	lb/hr
Total Centrate Solids	3,391	lb/day	141	lb/hr
Total Cake Solids	64,432	lb/day	32	ton/day
Total Wet Cake Solids	214,773	Wet lb/day	107	Wet ton/day
Total Cake Flow to Trucks	17.9	gpm	25,759	gpd
Cake Solids Concentration	300,000	mg/l	30%	%
Total Centrate Flow to Plant	60	gpm	86,239	gpd
Centrate Solids Concentration	4,716	mg/l	0.47%	%

OTHER

Centrate Surge Tank Size

Retention Time Required	**10**	min		
Centrate Flowrate	60	gpm	86,239	gpd
Total Volume Required	80	ft^3	599	gallons
Number of Tanks	**2**			
Volume Required per Tank	40	ft^3	299	gallons
Tank Height	**8**	ft		
Tank Diameter	2.5	ft		

Sulfuric Acid Tank Size

# of Days of Sulfuric Acid Supply Required	**7**	days		
Total Sulfuric Flowrate	0.06	gpm	89	gpd
Total Volume Required	83	ft^3	620	gallons
Tank Height	**10**	ft		
Tank Diameter	3.2	ft		

Influent Preheat Heat Exchanger

Heat Recovery per Shaft	1,587,574	Btu/hr		
Heat Transfer Coefficient	**75**	Btu/hr-°F-ft^2		
Sludge to Sludge Temperature Approach	**0**	C		
Hot Product Supply	60.0	C	140	F
Cool Influent Sludge	15.0	C	59	F
Tempered Product Return	**37.5**	C	99.5	F
Preheated Sludge to Shaft	37.5	C	99.5	F
Log Mean Differential Temperature	22.5	C	41	F
# of Heat Exchangers	2			
Surface Area for each Heat Exchanger	522	ft^2		
Flowrate of product sludge through exchanger	78	gpm		

Shaft Internal Recycle Heat Exchanger

Cooling Requirement per Shaft	3,124,456 Btu/hr	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Sludge to Water Temperature Approach	5 C	
Sludge Supply	60.0 C	140 F
Cooling Water Supply	10.0 C	50 F
Sludge Return	45.0 C	113 F
Cooling Water Return	50.0 C	122 F
Log Mean Differential Temperature	20.0 C	36 F
# of Heat Exchangers	2	
Surface Area for each Heat Exchanger	1158 ft^2	
Flowrate of reactor sludge through exchanger	174 gpm	
Flowrate of water through exchanger	87 gpm	

Biofilter Heat Exchanger

Cooling Requirement per Biofilter	1,115,025 Btu/hr	
Biofilter Temperature	30 C	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Water to Water Temperature Approach	0 C	
Biofilter Supply	30.0 C	86 F
Cooling Water Supply	10.0 C	50 F
Biofilter Return	20.0 C	68 F
Cooling Water Return	20.0 C	68 F
Differential Temperature	10.0 C	18 F
Surface Area for each Heat Exchanger	825 ft^2	
# of Heat Exchangers	2	
Flowrate of biofilter liquor through exchanger	124 gpm	
Flowrate of water through exchanger	124 gpm	
Turnover time in the biofilter	969 min	16.2 hrs

Energy Recovery and Generation

Total Heat Recovery from VerTad System	11,658,185 Btu/hr	
Heat Generation from VerTad Internal Recycle	6,251,097 Btu/hr	(As 122°F cooling water return)
Heat Removal from Biofilter	2,230,829 Btu/hr	
Heat used to Preheat Sludge to VerTad	3,176,259 Btu/hr	
Heat available from the Compressors	524,513 Btu/hr	

North Treatment Plant VERTAD Alternative Analysis

Table 4- : CHEMICAL COST ANALYSIS

Process Area	Anaerobic Base Case	Class A Anaerobic	Alt NP1 4 day / 2 D	Alt NP2 1.4 day / 3 D	Alt NP3 4 day VERTAD
VERTAD flotation	0	0	$ 16,210	$ 16,256	$ 16,210
Centrifuge Dewatering	$ 401,637	$ 373,544	$ 234,764	$ 255,525	$ 290,253
Total annual cost	$ 401,637	$ 373,544	$ 250,973	$ 271,781	$ 306,463

Polymer usage for centrifuge is 25% greater than BFPs				Polymer dosage	
Acid Cost is	$	100.00	**/ ton**	lb/dt	
Polymer cost is	$	1.80	**/ lb polymer**		100%
anaerobic	$	62.50	/dt of biosolids	35	
Vertad-anaer.	$	45.00	/dt of biosolids	25	1
Vertad	$	36.00	/dt of biosolids	20	1

Polymer Demand (tons/year)	112	104	65	71	81

North Treatment Plant VERTAD Alternative Analysis

Operation and maintenance Costs - Equipment Maintenance

				Operating Equipment					Annual Maintenance Cost				
Equipment Name	Equip Purchase Cost ($)	Annual Mainten. percent (%)	Annual Mainten. Cost $/yr	Anaerobic Base Case w/ Cent. No.	Class A Anaerobic No.	Alt NP1 4 day / 2 D w/ Cent. No.	Alt NP2 1.4 day / 3 D w/ Cent. No.	Alt NP3 4 day VERTAD Class A No.	Anaerobic Base Case w/ Cent. $/yr	Class A Anaerobic $/yr	Alt NP1 4 day / 2 D w/ Cent. $/yr	Alt NP2 1.4 day / 3 D w/ Cent. $/yr	Alt NP3 4 day VERTAD w/ Cent. $/yr
Blending Tank Equipment													
Blending tank circ pump	35,000	5%	1,750	1	1	1	1	1	1,750	1,750	1,750	1,750	1,750
Digester feed pump	25,000	10%	2,500	2	2	2	3	0	5,000	5,000	5,000	7,500	-
Digester withdrawal pumps	25,000	10%	2,500	2	3	2	3	0	5,000	7,500	5,000	7,500	-
Digester Equipment													
Digester mixing compressors	87,000	5%	4,350	2	3	2	2	0	8,700	13,050	8,700	8,700	-
Grinders	15,000	10%	1,500	4	9	4	4	0	6,000	13,500	6,000	6,000	-
Circ sludge pump 1	20,000	5%	1,000	2	3	2	2	0	2,000	3,000	2,000	2,000	-
Circ sludge pump 2 (hex)	20,000	5%	1,000	2	3	2	2	0	2,000	3,000	2,000	2,000	-
Heating System Equipment													
Heat extractors	350,000	5%	17,500	2	2	0	0	0	35,000	35,000	-	-	-
HWRS pump	20,000	5%	1,000	2	6	2	2	0	2,000	6,000	2,000	2,000	-
Boilers	60,000	10%	6,000	0	2	0	0	0	-	12,000	-	-	-
VERTAD Equipment													
Supply pumps	25,000	10%	2,500	0	0	2	2	2	-	-	5,000	5,000	5,000
Anaerobic feed pumps	25,000	10%	2,500	0	0	2	2	2	-	-	5,000	5,000	5,000
Product Tank Mixers	5,000	10%	500	0	0	2	1	1	-	-	1,000	500	500
Compressors - 350 hp	150,000	5%	7,500	0	0	2	0	0	-	-	15,000	-	-
175 hp	35,000	5%	1,750	0	0	0	2	0	-	-	-	3,500	-
532 hp	120,000	5%	6,000	0	0	0	0	2	-	-	-	-	12,000
Biofilter pumps	150,000	2%	3,000	0	0	2	2	2	-	-	6,000	6,000	6,000
Flotation Tank Equipment													
Acid pumps	15,000	10%	1,500	0	0	1	1	1	-	-	1,500	1,500	1,500
Subnatant pump	25,000	5%	1,250	0	0	2	2	2	-	-	2,500	2,500	2,500
Scraper	20,000	5%	1,000	0	0	4	4	4	-	-	4,000	4,000	4,000
Heat Exchangers	25,000	10%	2,500	0	0	6	4	6	-	-	15,000	10,000	15,000
Hold Tank Equipment													
Grinder	15000	10%	1,500	0	1	0	0	0	-	1,500	-	-	-
Circ sludge pump	25000	5%	1,250	0	1	0	0	0	-	1,250	-	-	-
HWRS Pump	20000	5%	1,000	0	1	0	0	0	-	1,000	-	-	-
Tank Withdrawal pump	25000	10%	2,500	0	3	0	0	0	-	7,500	-	-	-
Automatic Valves	32000	10%	3,200	0	1	0	0	0	-	3,200	-	-	-
Subtotal (Digestion)									67,450	114,250	87,450	75,450	53,250
Dewatering Equipment													
Centrifuge	250,000	3%	7,500	2	2	2	2	2	15,000	15,000	15,000	15,000	15,000
Centrifuge feed pump	25,000	10%	2,500	4	4	4	4	4	10,000	10,000	10,000	10,000	10,000
Polymer feed pump	7,500	10%	750	4	4	4	4	4	3,000	3,000	3,000	3,000	3,000
Centridry	1,000,000	3%	30,000	0	0	0	0	0	-	-	-	-	-
Subtotal (Dewatering)									28,000	28,000	28,000	28,000	28,000
Total Annual Maintenance Cost									$ 95,450	$ 142,250	$ 115,450	$ 103,450	$ 81,250

Table 4- : Operation and maintenance Costs - Labor Costs

				Anaerobic	Class A Anaerobic	Alt NP1	Alt NP2	Alt NP3
Operating Labor for Digestion								
Digester Labor (1995)	$	262,793						
Inflate to 1999 dollars	$	295,776						
Divide by 5.5 tanks	$	53,777						
(Strategy is that each digester is a tank, the BST is a tank, each VERTAD reactor is 3/4 tank, each flotation tank is 1/2 tank and each Blending tank is 1/2 a tank)								
			No of tanks	3.5	5.5	6	6	3.5
			Annual Labor	$ 188,221	$ 295,776	$ 322,665	$ 322,665	$ 188,221
Operating Labor for Centrifuge Dewatering						1	VERTAD labor multiplier	
Estimated Operating Cost per			No. of Centrifuges's	1	1	1	1	2
Centrifuge installed	$	90,000	Annual Labor	$ 90,000	$ 90,000	$ 90,000	$ 90,000	$ 180,000

North Treatment Plant VERTAD Alternative Analysis

Table 4- : Energy Cost Analysis - Digester Gas and Hot Water

Heating value of gas	BTU/cft	600
Net value of digester gas	$/therm	$ 0.07
Net value of hot water	$/therm	$ 0.02
Average boiler efficiency	percent	78%

100%

Average Annual Values	Units	Anaerobic Base Case	Class A Anaerobic	Alt NP1 4 day / 2 D	Alt NP2 1.4 day / 3 D	Alt NP3 4 day VERTAD
Number Digesters		2		2	2	0
Digester Gas Production	cft/day	438,758	518,532	217,339	401,821	0
Heating Value @ 600 BTU/cft	BTU/day	263,254,860	311,119,380	130,403,687	241,092,383	-
Heating Value	Therms/yr	960,880	1,135,586	475,973	879,987	-
Heat Required for Digestion	BTU/hr	2,665,591	3,165,094	2,332,292	2,332,292	0
Hot water from VERTAD	BTU/hr	-	-	(7,526,455)	(2,829,019)	(7,669,858)
Total Heat Required	BTU/hr	2,665,591	3,165,094	(5,194,163)	(496,726)	(7,669,858)
Excess Heat Available	BTU/hr	-	-	5,194,163	496,726	7,669,858
	Therms/yr	-	-	455,009	43,513	671,880
Gas Demand for digesters	BTU/hr	3,417,424	4,057,813	-	-	-
Gas Demand for digesters	Therms/yr	299,366	355,464	-	-	-
Excess Gas Available for Sale	Therms/yr	661,514	780,121	475,973	879,987	-
Annual Revenue from Gas Sold	$/yr	$ 46,306	$ 54,608	$ 33,318	$ 61,599	$ -
Annual value of excess hot water	$/yr	$ -	$ -	$ 9,100	$ 870	$ 13,438

- Anaerobic Digester Heat Demand is generated by the heat extractors and accounted for as part of electrical demand

North Treatment Plant VERTAD Alternative Analysis

Table 4- : Electrical Costs

Equipment Name	HP (name plate)	HP draw (est)	Operating Equipment Alt 1 No Proj No.	Operating Equipment Class A anaerobic No.	Operating Equipment Alt NP1 4 day / 2 D No.	Operating Equipment Alt NP2 1.4 day / 3 D No.	Operating Equipment Alt NP3 4 day VERTAD Class A	Constant Electricity Usage Alt 1B No Project HP draw	Constant Electricity Usage Class A anaerobic HP draw	Constant Electricity Usage Alt NP1 4 day / 2 D HP draw	Constant Electricity Usage Alt NP2 1.4 day / 3 D HP draw	Constant Electricity Usage Alt NP3 4 day VERTAD HP draw	Variable Electricity Usage Alt 1B No Project HP draw	Variable Electricity Usage Class A anaerobic HP draw	Variable Electricity Usage Alt NP1 4 day / 2 D HP draw	Variable Electricity Usage Alt NP2 1.4 day / 3 D HP draw	Variable Electricity Usage Alt NP3 4 day VERTAD HP draw
Blending Tank Equipment																	
Blending Tank Circ pump	20	18	1	1	1	1	1	18	18	18	18	18					
Digester Feed pump	30	28	2	2	2	2	0						56	56	56	56	0
Digester Withdrawal Pumps	15	13	2	3	2	2	0						26	39	26	26	0
Digester Equipment (number digesters)			2		2	2	0										
Digester Mixing Compressors	35	33	2	3	2	2	0	66	99	66	66	0					
Grinders	10	8	4	9	4	4	4	32	72	32	32	32					
Circ sludge pump 1	35	33	2	3	2	2	2	66	99	66	66	66					
Circ sludge pump 2 (hex)	30	28	2	3	2	2	2	56	84	56	56	56					
Heating System Equipment																	
Heat extractors																	
Hp/MMBTUH	68	68															
MMBTUH required			7.3	0	0	0	0										
HP draw													0	0	0	0	0
HWRS pump	20	18	4	6	2	2	0	72	108	36	36	0					
VERTAD Equipment (Reactors)			0	0	2	2	2										
Supply Pumps	2	1.8	0	0	2	2	2	0	0	3.6	3.6	3.6					
Anaerobic feed pumps	1	1	0	0	2	2	0	0	0				0	0	2	2	0
Product Tank Mixer	10	9	0	0	2	2	1	0	0	18	18	9					
Compressors	350	340	0	0	2			0	0				0	0	678	0	0
	175	170				2										339	
	530	345					2										678
Biofilter pumps	6	5.7	0	0	2	2	2	0	0	11.4	11.4	11.4					
Flotation Equipment																	
Acid Pump	0.5	0.5	0	0	1	1	1			0.5	0.5	0.5					
Subnatant Pump	2	2	0	0	2	2	2			4	4	4					
Scraper	1	1	0	0	4	4	4			4	4	4					
Heat exchangers	0	0	0	0	6	4	6			0	0	0					
Hold Tank Equipment																	
Grinder	10	8		1				0	8	0	0	0	0	0	0	0	0
Circ sludge pump	50	40		1				0	40	0	0	0	0	0	0	0	0
HWRS Pump	30	20		1				0	20	0	0	0	0	0	0	0	0
Tank Withdrawal pump	25	15		3				0	45	0	0	0	0	0	0	0	0
Total Hp Digestion								310	593	316	316	205	82	95	762	423	678
Total electricity draw (kW)								231	442	235	235	152	61	71	568	315	505
Total cost/yr (Digestion)								$ 92,341	$ 176,639	$ 93,979	$ 93,979	$ 60,915	$ 24,426	$ 28,298	$ 226,908	$ 125,907	$ 201,887
Dewatering Equipment																	
Centrifuges main drive	125	100	0.6	0.5	0.5	0.5	1.1						63	55	48	52	112
Centrifuges back drive	30	18	0.6	0.5	0.5	0.5	1.1						11	10	9	9	20
Centridry	638	477	0.0	0.0	0.0	0.0	0.0						0	0	0	0	0
Centrifuge feed sludge pump	30	20	0.6	0.5	0.5	0.5	1.1						13	11	10	10	22
Polymer feed pump	2	1	0.6	0.5	0.5	0.5	1.1						1	1	0	1	1
Conveyors	5	3	2.0	2.0	2.0	2.0	2.0	6	6	6	6	6					
Total Hp Dewatering								6	6	6	6	6	88	76	66	72	156
Total electricity draw (kW)								4	4	4	4	4	66	57	49	54	116
Total cost/yr Dewatering								$ 1,787	$ 1,787	$ 1,787	$ 1,787	$ 1,787	$ 26,271	$ 22,601	$ 19,727	$ 21,471	$ 46,340
Cost of Electricity	0.0456	$/kW-hr	100%														

VERTAD Evaluation

CAPITAL AND ANNUAL COST ANALYSIS

Table 4- : Capital Cost Estimate

Digestion

Category		Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Site Work		0		50,000	50,000	50,000
Demolition		0				
Anaerobic Digestion		6,500,126	6,807,171	3,269,558	3,805,490	-
Structures		0	0	560,000	390,000	700,000
Equipment & Mech		0	0	640,000	540,000	700,000
Patent Fee		0	0	-	-	-
Electrical/ I&C		0	0	230,000	220,000	220,000
Testing/ Start-up		0	0	incl	incl	incl
NORAM provided equipment and engineering				3,570,000	2,890,000	3,920,000
Cased Reactors	1			3,260,000	1,620,000	4,490,000
Subtotal		6,500,126	6,807,171	11,579,558	9,515,490	10,080,000
Contractor Indirects, OH@P	35%	2,275,044	2,382,510	1,161,845	1,349,421	17,500
NORAM estimate	11.6%			544,040	321,320	708,760
NORAM provided	5%			178,500	144,500	196,000
Subtotal		8,775,170	9,189,681	13,463,944	11,330,731	11,002,260
Contingency	30%	2,632,551	2,756,904	2,314,053	2,436,570	848,831
Drilling	10%			379,051	192,904	490,081
NORAM provided	5%			207,548	172,068	213,933
Subtotal		11,407,721	11,946,585	16,364,596	14,132,272	12,555,105
Sales tax	8.4%	958,249	1,003,513	1,374,626	1,187,111	1,054,629
Subtotal		12,365,970	12,950,098	17,739,222	15,319,382	13,609,733
Allied Cost (35%)	35%	4,328,089	4,532,534	6,208,728	5,361,784	4,763,407
Less NORAM engineering	15%	-	-	(931,309)	(804,268)	(714,511)
Total		16,694,059	17,482,633	23,016,641	19,876,899	17,658,629
Anaerobic Digestion Capacity (gallons)		4,463,477	3,740,010	1,828,514	2,226,951	-
Digester volume Ratio design / ESRP Digester 5		1.63	1.37	0.67	0.82	-
Capital Cost Adjustment Factor		0.77	0.77	0.77	0.77	0.77
Multiple reactor multiplier			1.2			

Dewatering

Category		Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Site Work					-	
Demolition						
Structures		910,000	910,000	910,000	910,000	1,365,000
Equipment & Mech		1,562,460	1,562,460	1,562,460	1,562,460	2,135,006
Patent Fee						
Electrical/ I&C						
Testing/ Start-up						
Subtotal		2,472,460	2,472,460	2,472,460	2,472,460	3,500,006
Contractor Indirects, OH@P	35%	865,361	865,361	865,361	865,361	1,225,002
Subtotal		3,337,821	3,337,821	3,337,821	3,337,821	4,725,009
Contingency	30%	1,001,346	1,001,346	1,001,346	1,001,346	1,417,503
Subtotal		4,339,167	4,339,167	4,339,167	4,339,167	6,142,511
Sales tax	8.4%	364,490	364,490	364,490	364,490	515,971
Subtotal		4,703,657	4,703,657	4,703,657	4,703,657	6,658,482
Allied Cost (35%)	35%	1,646,280	1,646,280	1,646,280	1,646,280	2,330,469
Total		6,349,937	6,349,937	6,349,937	6,349,937	8,988,951
number units		2	2	2	2	3
		1				
		$ 9,374,000	base w/	0.77	cost adjust factor	
		0.335	*solids ratio*	*$ 4,038,401*	*adjusted cent cost*	

Total Capital Cost

	Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Grand Total Capital Expendature	$ 23,043,997	$ 23,832,570	$ 29,366,579	$ 26,226,836	$ 26,647,580

Table 4- : Annual Costs at Design Year Loading, 2019 (todays dollars)

Digestion

Category	Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Equipment Maintenance	$ 67,450	$ 114,250	$ 87,450	$ 75,450	$ 53,250
Operations Labor	$ 188,221	$ 295,776	$ 322,665	$ 322,665	$ 188,221
Power					
Fixed	$ 92,341	$ 176,639	$ 93,979	$ 93,979	$ 60,915
Variable	$ 24,426	$ 28,298	$ 226,908	$ 125,907	$ 201,887

Chemicals (acid)	$ -	$ -	$ 16,210	$ 16,256	$ 16,210
Hot water avoided cost	$ -	$ -	$ (9,100)	$ (870)	$ (13,438)
Gas Sale Net Revenue	$ (46,306)	$ (54,608)	$ (33,318)	$ (61,599)	$ -
Total Annual Cost	$ 326,132	$ 560,355	$ 704,794	$ 571,787	$ 507,045

Dewatering

Category	Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Equipment Maintenance	$ 28,000	$ 28,000	$ 28,000	$ 28,000	$ 28,000
Operations Labor	$ 90,000	$ 90,000	$ 90,000	$ 90,000	$ 180,000
Power					
Fixed	$ 1,787	$ 1,787	$ 1,787	$ 1,787	$ 1,787
Variable	$ 26,271	$ 22,601	$ 19,727	$ 21,471	$ 46,340
Chemicals (polymer)	$ 401,637	$ 373,544	$ 234,764	$ 255,525	$ 290,253
Total Annual Cost	$ 547,695	$ 515,932	$ 374,278	$ 398,783	$ 546,380

Biosolids Haul and Application

Category		Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Biosolids Haul and Application		$ 768,995	$ 673,072	$ 467,197	$ 508,513	$ 722,032
Wet Tons	1	27,464	24,038	16,686	18,161	25,787
$ / WT	100%	$ 28.00	$ 28.00	$ 28.00	$ 28.00	$ 28.00
Dry Tons	1	6,427	5,529	5,006	5,448	7,736
$ / DT		$ 119.66	$ 121.74	$ 93.33	$ 93.33	$ 93.33

Total Annual Cost (Year 2019, Todays Dollars)

Category	Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Total Annual Cost	$ 1,642,822	$ 1,749,360	$ 1,546,268	$ 1,477,083	$ 1,775,457
Annual Costs expected to not vary with flow	$ 467,799	$ 706,453	$ 623,881	$ 611,881	$ 512,174
Annual Costs expected to vary with flow	$ 1,175,022	$ 1,042,907	$ 922,387	$ 865,202	$ 1,263,284

Table 4- : Annual Costs by Year *

Year	Annual Average Flow	Percentage of Year 2019 flow	Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
2010	18.0	50.0%	1,055,310	1,227,906	1,085,075	1,044,482	1,143,816
2011	19.1	53.1%	1,092,030	1,260,497	1,113,899	1,071,520	1,183,293
2012	20.3	56.3%	1,128,749	1,293,088	1,142,724	1,098,557	1,222,771
2013	21.4	59.4%	1,165,469	1,325,679	1,171,549	1,125,595	1,262,248
2014	22.5	62.5%	1,202,188	1,358,269	1,200,373	1,152,632	1,301,726
2015	23.6	65.6%	1,238,908	1,390,860	1,229,198	1,179,670	1,341,204
2016	24.8	68.8%	1,275,627	1,423,451	1,258,022	1,206,708	1,380,681
2017	25.9	71.9%	1,312,347	1,456,042	1,286,847	1,233,745	1,420,159
2018	27.0	75.0%	1,349,066	1,488,633	1,315,672	1,260,783	1,459,637
2019	28.1	78.1%	1,385,785	1,521,224	1,344,496	1,287,820	1,499,114
2020	29.3	81.3%	1,422,505	1,553,814	1,373,321	1,314,858	1,538,592
2021	30.4	84.4%	1,459,224	1,586,405	1,402,145	1,341,895	1,578,069
2022	31.5	87.5%	1,495,944	1,618,996	1,430,970	1,368,933	1,617,547
2023	32.6	90.6%	1,532,663	1,651,587	1,459,795	1,395,970	1,657,025
2024	33.8	93.8%	1,569,383	1,684,178	1,488,619	1,423,008	1,696,502
2025	34.9	96.9%	1,606,102	1,716,769	1,517,444	1,450,046	1,735,980
2026	36.0	100.0%	1,642,822	1,749,360	1,546,268	1,477,083	1,775,457

* Annual cost for each year is calculated as the sum of the portion of the annual cost not expected to vary with flow and the prorated annual cost expected to vary with flow.

1.125

Table 4- : Avoided Costs

	Anaerobic Base	*Anaerobic Class A*	*Alternative NP1*	*Alternative NP2*	*Alternative NP3*
Digester					
Year in which a new digester is required	2017	2017	2017	2017	2017
Years away	18	18	18	18	18
Inflated cost	$ 7,580,934	$ 7,580,934	$ 7,580,934	$ 7,580,934	$ 7,580,934
Present Worth	$ 2,655,933	$ 2,655,933	$ 2,655,933	$ 2,655,933	$ 2,655,933
Off-set Capital Cost	$ -	$ -	$ -	$ -	$ -

Capital cost of a digester in 1999 dollars	$ 4,453,000
Capacity of Digester (gallons)	2,731,095
Capital cost for 2 belt filter presses including Bldg expansion 1999 dollars	3,013,000

0% surcharge for control building
not included in Digester 5 costs

PRESENT WORTH COST ANALYSIS -Life Cycle Costs

		Anaerobic Base	Anaerobic Class A	Alternative NP1	Alternative NP2	Alternative NP3
Inflated Annual Costs Year	inflation years					
2003	4	$ 1,187,761	$ 1,382,019	$ 1,221,261	$ 1,175,574	$ 1,287,374
2004	5	$ 1,265,962	$ 1,461,261	$ 1,291,315	$ 1,242,185	$ 1,371,761
2005	6	$ 1,347,786	$ 1,544,014	$ 1,364,472	$ 1,311,735	$ 1,460,052
2006	7	$ 1,433,380	$ 1,630,417	$ 1,440,857	$ 1,384,340	$ 1,552,406
2007	8	$ 1,522,896	$ 1,720,615	$ 1,520,597	$ 1,460,120	$ 1,648,988
2008	9	$ 1,616,494	$ 1,814,757	$ 1,603,824	$ 1,539,202	$ 1,749,967
2009	10	$ 1,714,336	$ 1,912,999	$ 1,690,677	$ 1,621,714	$ 1,855,520
2010	11	$ 1,816,595	$ 2,015,503	$ 1,781,297	$ 1,707,792	$ 1,965,832
2011	12	$ 1,923,446	$ 2,122,434	$ 1,875,833	$ 1,797,575	$ 2,081,093
2012	13	$ 2,035,073	$ 2,233,968	$ 1,974,438	$ 1,891,207	$ 2,201,500
2013	14	$ 2,151,666	$ 2,350,284	$ 2,077,271	$ 1,988,840	$ 2,327,258
2014	15	$ 2,273,424	$ 2,471,568	$ 2,184,497	$ 2,090,629	$ 2,458,581
2015	16	$ 2,400,551	$ 2,598,014	$ 2,296,287	$ 2,196,735	$ 2,595,688
2016	17	$ 2,533,259	$ 2,729,822	$ 2,412,818	$ 2,307,326	$ 2,738,809
2017	18	$ 2,671,769	$ 2,867,200	$ 2,534,275	$ 2,422,576	$ 2,888,182
2018	19	$ 2,816,310	$ 3,010,364	$ 2,660,847	$ 2,542,664	$ 3,044,051
2019	20	$ 2,967,119	$ 3,159,538	$ 2,792,733	$ 2,667,777	$ 3,206,674
PW Annual Costs		$ 15,193,235	$ 16,827,481	$ 14,872,170	$ 14,256,316	$ 16,440,464
PW Capital Costs		$ 23,043,997	$ 23,832,570	$ 29,366,579	$ 26,226,836	$ 26,647,580
Subtotal		$ 38,237,232	$ 40,660,051	$ 44,238,748	$ 40,483,152	$ 43,088,044
PW of Avoided Capital Costs		$ -	$ -	$ -	$ -	$ -
Total Present Worth		$ 38,237,000	$ 40,660,000	$ 44,239,000	$ 40,483,000	$ 43,088,000

APPENDIX D (b)

South Treatment Plant Cost Estimates

HIGH END ASSUMPTIONS COST MODEL

COMMON TO ALL ALTERNATIVES

Table 1.1
Assumptions Common to all Alternatives

Planning Period	yr	2010
Influent Waste Water Flow	mgd	104
Peaking Factors		
Annual Average		1
Peak 3-week		1.22
Peak Week		1.52
Peak Day		1.66
Solids Unit Production Rate		
Primary Sludge	lbs-tss/MG	1310
Waste Activated Sludge	lbs-tss/MG	770
Thickened Sludge	lbs-tss/MG	1870
Solids Concentration		
Primary Sludge	percent solids	0.6%
Waste Activated Sludge	percent solids	0.5%
Thickened Sludge	percent solids	5.5%
Volatile Solids Ratio		
Primary Sludge	percent of tss	80%
Waste Activated Sludge	percent of tss	76%
Thickened Sludge	percent of tss	79%
Summer Sludge Temperature	F	71
Winter Sludge Temperature	F	57
Thermophilic Operating Temperature	F	135
Mesophilic Operating Temperature	F	95
Digester Cone Volume Percent Active	percent	75%
Digester Gas Production	(cf/lb vs dest)	15

Table 1.2
Projected Solids Process Flows and Loads

Raw Solids	Flow	Flow	TSS	VSS
Primary Sludge	gpm	gpd	lbs/day	lbs/day
Average Annual	1,891	2,722,622	136,240	108,992
Peak 3-Week	2,304	3,317,346	166,000	132,800
Peak Day	3,136	4,516,387	226,000	180,800
Waste Activated Sludge				
Average Annual	1,334	1,920,384	80,080	60,861
Peak 3-Week	1,627	2,342,868	97,698	74,250
Peak Day	2,214	3,187,837	132,933	101,029
Mixed Sludge				
Average Annual	3,224	4,643,006	216,320	169,853

Peak 3-Week		3,931	5,660,214	263,698	207,050
Peak Day		5,350	7,704,224	358,933	281,829
Thickening					
DAFTS					
Number of Units	no.	4			
Diameter	ft	55			
Number of Units	no.	2			
Diameter	ft	65			
Total Surface Area	sft	16,140			
Solids Loading, All DAFT's in service					
Average Annual	lb/sf/day	13.4			
Peak 3-Week	lb/sf/day	16.3			
Peak Day	lb/sf/day	22.2			
Solids Loading, one DAFT out of service					
Average Annual	lb/sf/day	16.9			
Peak 3-Week	lb/sf/day	20.6			
Peak Day	lb/sf/day	28.0			
Thickened Solids		Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Thickened Sludge					
Average Annual		294	423,981	194,480	153,639
Peak 3-Week		359	517,257	237,266	187,440
Peak Day		489	703,808	322,837	255,041
Thickened Sludge Blending Tank					
Number of Units	no.	1			
Diameter	ft	8			
Sidewater Depth	ft	15			
Tank Volume	gallons	5,640			
Total Volume	gallons	5,640			
Detention Time					
Average Annual	minutes	19.2			
Peak 3-Week	minutes	15.7			
Peak Day	minutes	11.5			

ALTERNATIVE 3B - THERMO-MESO DIGESTION AND CONVERT TO CENTRIFUGE DEWATERING

Table 4.1
Thermo-Meso Digestion Assumptions

Total Volatile Solids Reduction in Digesters	percent	65.00%
Assumed contribution to total digestion		
Themophilic Phase	fraction	0.65
Mesophilic Phase	fraction	0.35
Dewatering Solids Capture	percent	92.5%
Dewatered Cake Solids	pecent solids	25.00%

Table 4.2
Thermo-Meso Digestion Flows and Loads

Anaerobic Thermophillic Digester					
Existing					
Number of Units (existing)	no.	1			
Diameter	ft	100			
Sidewater Depth	ft	44.5			
Tank Cylindrical Volume	gallons	2,614,276			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75% of con	gallons	2,789,842			
Total Existing Volume	gallons	2,789,842			
Total Volume	gallons	2,789,842			
Detention Time					
Average Annual	days	6.6			
Peak 3-Week	days	5.4			
Peak Day	days	4.0			
Volatile Solids Loading					
Average Annual	lbs-vs/cf/day	0.41			
Peak 3-Week	lbs-vs/cf/day	0.50			
Peak Day	lbs-vs/cf/day	0.68			
Heat Demand					
Digester Heat Loss to Control Bldg	BTU/hr	101,531			
Digester Heat Loss from Shell of 1 Digester	BTU/hr	495,907			
Average Annual	BTU/hr	12,080,252			
Peak 3-Week	BTU/hr	14,606,471			
Peak Day	BTU/hr	19,658,909			
Thermophilic Gas Production					
Average Annual	cf/day	973,688			
Peak 3-Week	cf/day	1,187,900			
Peak Day	cf/day	1,616,323			
Thermophilic Digested Solids		Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Digested Sludge					
Average Annual		294	423,981	135,174	92,566
Peak 3-Week		359	517,257	164,912	112,930
Peak Day		489	703,808	224,388	153,659
Anaerobic Mesophillic Digesters					
Existing					
Number of Units (existing)	no.	3			
Diameter	ft	100			
Sidewater Depth	ft	43.5			
Tank Cylindrical Volume	gallons	2,555,529			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75% of con	gallons	2,731,095			

Total Existing Volume	gallons	8,193,284			
Total Volume	gallons	8,193,284			
Detention Time					
All Digesters in Service					
Average Annual	days	19.3			
Peak 3-Week	days	15.8			
Peak Day	days	11.6			
One Digester Out of Service, BST as a Digester					
Average Annual	days	18.4			
Peak 3-Week	days	15.1			
Peak Day	days	11.1			
Volatile Solids Loading					
All Digesters in Service					
Average Annual	bs-vs/cf/day	0.08			
Peak 3-Week	bs-vs/cf/day	0.10			
Peak Day	bs-vs/cf/day	0.14			
One Digester Out of Service					
Average Annual	bs-vs/cf/day	0.09			
Peak 3-Week	bs-vs/cf/day	0.11			
Peak Day	bs-vs/cf/day	0.15			
Cooling Demand					
Digester Heat Loss to Control Bldg	BTU/hr	142,144	Per Digester		
Digester Heat Loss from Shell	BTU/hr	179,445			
Average Annual	BTU/hr	-5,208,143	-1,736,048		
Peak 3-Week	BTU/hr	-6,503,640	-2,167,880		
Peak Day @ 135 F	BTU/hr	-9,094,634	-3,031,545		
Peak Day @ 140 F	BTU/hr	-10,316,523	-3,438,841		
Gas Production Meso Digestion					
Average Annual	cf/day	524,294			
Peak 3-Week	cf/day	639,638			
Peak Day	cf/day	870,328			
Digester Gas Production Total Digestion Process					
Average Annual	cf/day	1,497,982			
Peak 3-Week	cf/day	1,827,538			
Peak Day	cf/day	2,486,650			
Digested Solids		Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Digested Sludge					
Average Annual		294	423,981	94,615	53,774
Peak 3-Week		359	517,257	115,430	65,604
Peak Day		489	703,808	157,060	89,264
Digested Sludge Storage					
Blending Storage Tank					
Diameter	ft	100			
Sidewater Depth	ft	36.75			
Tank Cylindrical Volume	gallons	2,158,981			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75% of con	gallons	2,334,547			
Total BST Volume	gallons	2,334,547			
Maximum Storage Capacity (no cone volume considered)					
Average Annual	days	5.1			
Peak 3-Week	days	4.2			
Peak Day	days	3.1			
Dewatering		Flow gpm	TSS lbs/day	Concentration % solids	
Loading					
Average Annual		294	94,615	2.68%	
Max 3-Week		359	115,430	2.68%	
Max Week	⇨	448	143,814	2.68%	
Peak Day		489	157,060	2.68%	
Centrifuge Dewatering					

Number of Units Needed (Max Week + 1 unit	no.	3.4
Nominal Size	meters	CP3074
Dewatering Capacity/Centrifuge	dry lbs/hr	2,500
Operating Units at Annual Average Condition	no.	1.6
Operating Units at Max 3-week Condition	no.	1.9
Operating Units at Max Week Condition	no.	2.4
Operating Units at Peak Day Condition	no.	2.6
Dewatered Biosolids		
Dry Solids TSS (lbs)		
Average Annual	lbs/day	87,518
Peak 3-Week	lbs/day	106,772
Peak Day	lbs/day	145,281
Dry Solids TSS (tons)		
Average Annual		44
Peak 3-Week		53
Peak Week		67
Peak Day		73
Dry Solids VSS (lbs)		
Average Annual	lbs/day	49,741
Peak 3-Week	lbs/day	60,684
Peak Day	lbs/day	82,570
Wet Cake (lbs)		
Average Annual	wet lbs/day	350,074
Peak 3-Week	wet lbs/day	427,090
Peak Day	wet lbs/day	581,122
Wet Cake (tons)		
Average Annual	wet ton/day	175
Peak 3-Week	wet ton/day	214
Peak Day	wet ton/day	291

Table 4.3
Thermo-Meso Digestion Capacity Extension Analysis

Thermophilic Digesters			
Evaluation Criteria			
1 Peak Day Thermophilic HRT >,= 3.5 days			
Volume	gal	2,789,842	
Max allowable Peak Day flow	gpd	797,098	
Max allowable average sludge flow =		480,179	governs
Convert to Annual Average influent	mgd	116	
Year ESRP to Reach this flow	YR	2024	
Mesophilic Digesters			
2 Peak Day flow w/ all Digesters in service, DT >, = 10 days			
Total Volume	gal	8,193,284	
Acceptable peak sludge flow =	gpd	819,328	
therefore average flow =	gpd	493,571	
3 Max 3-week flow w/ 1 Digester out of service & the BST as Meso Digester, DT >,= 10			
Total Volume	gal	7,796,736	
Acceptable peak sludge flow =	gpd	779,674	
therefore average flow =	gpd	639,076.73	

ALTERNATIVE 6 modified - THERMO-MESO DIGESTION W/ HOLD TANKS, *0 CENTRIDRY*, 3 CENTRIFUGES

CLASS A Process

Table 6.1
Thermo-Meso Class A Digestion Assumptions

Total Volatile Solids Reduction in Digesters	percent	65.00%
Assumed contribution to total digestion		
Themophilic Phase	fraction	0.65
Mesophilic Phase	fraction	0.35
Dewatering Solids Capture	percent	92.5%
Centrifuge Dewatered Cake Solids	percent solids	25.0%
Centridry Dried Cake Solids	percent solids	55.0%
Dryed Product Capacity	lbs/hr	0

Table 6.2
Thermo-Meso Class A Digestion Flows and Loads

Anaerobic Thermophillic Digester					
Existing					
Number of Units (existing)	no.	1			
Diameter	ft	100			
Sidewater Depth	ft	44.5			
Tank Cylindrical Volume	gallons	2,614,276			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75	gallons	2,789,842			
Total Existing Volume	gallons	2,789,842			
Total Volume	gallons	2,789,842			
Detention Time					
Average Annual	days	6.6			
Peak 3-Week	days	5.4			
Peak Day	days	4.0			
Volatile Solids Loading					
Average Annual	lbs-vs/cf/day	0.41			
Peak 3-Week	lbs-vs/cf/day	0.50			
Peak Day	lbs-vs/cf/day	0.68			
Heat Demand					
Digester Heat Loss to Control Bldg	BTU/hr	101,531			
Digester Heat Loss from Shell of 1 I	BTU/hr	495,907			
Average Annual	BTU/hr	12,080,252			
Peak 3-Week	BTU/hr	14,606,471			
Peak Day	BTU/hr	19,658,909			
Thermophilic Gas Production					
Average Annual	cf/day	973,688			
Peak 3-Week	cf/day	1,187,900			
Peak Day	cf/day	1,616,323			
Thermophilic Digested Solids		Flow	Flow	TSS	VSS
Digested Sludge		gpm	gpd	lbs/day	lbs/day
Average Annual		294	423,981	129,567	88,727
Peak 3-Week		359	517,257	158,072	108,246
Peak Day		489	703,808	215,082	147,286
Class A Hold Tanks					
Temp min temp for hold tank	F	135			
Temp min temp for hold tank	C	57.22			
Hold Time	hrs	11.7			

Size of 1 Hold Tank (at peak day)	gal	343,495			
Rough Dimensions	diaxht	45x31			
Estimated Heat Demand	BTU/hr	400,000			
Anaerobic Mesophillic Digesters					
Existing					
Number of Units (existing)	no.	3			
Diameter	ft	100			
Sidewater Depth	ft	41.0			
Tank Cylindrical Volume	gallons	2,408,659			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75	gallons	2,584,225			
Total Existing Volume	gallons	7,752,675			
Total Volume	gallons	7,752,675			
Detention Time					
All Digesters in Service					
Average Annual	days	18.3			
Peak 3-Week	days	15.0			
Peak Day	days	11.0			
One Digester Out of Service, BST as a Digester					
Average Annual	days	17.7			
Peak 3-Week	days	14.5			
Peak Day	days	10.7			
Volatile Solids Loading					
All Digesters in Service					
Average Annual	lbs-vs/cf/day	0.09			
Peak 3-Week	lbs-vs/cf/day	0.10			
Peak Day	lbs-vs/cf/day	0.14			
One Digester Out of Service					
Average Annual	lbs-vs/cf/day	0.09			
Peak 3-Week	lbs-vs/cf/day	0.11			
Peak Day	lbs-vs/cf/day	0.15			
Cooling Demand					
Digester Heat Loss to Control Bldg	BTU/hr	142,144	Per Digester		
Digester Heat Loss from Shell	BTU/hr	179,445			
Average Annual	BTU/hr	-5,208,143	-1,736,048		
Peak 3-Week	BTU/hr	-6,503,640	-2,167,880		
Peak Day @ 135 F	BTU/hr	-9,094,634	-3,031,545		
Gas Production Meso Digestion					
Average Annual	cf/day	524,294			
Peak 3-Week	cf/day	639,638			
Peak Day	cf/day	870,328			
Digester Gas Production Total Digestion Process					
Average Annual	cf/day	1,497,982			
Peak 3-Week	cf/day	1,827,538			
Peak Day	cf/day	2,486,650			
Digested Solids		Flow gpm	Flow gpd	TSS lbs/day	VSS lbs/day
Digested Sludge					
Average Annual		294	423,981	94,615	53,774
Peak 3-Week		359	517,257	115,430	65,604
Peak Day		489	703,808	157,060	89,264
Digested Sludge Storage					
Blending Storage Tank					
Diameter	ft	100			
Sidewater Depth	ft	36.75			
Tank Cylindrical Volume	gallons	2,158,981			
Tank Cone Volume	gallons	234,088			
Effective Tank Volume (includes 75	gallons	2,334,547			
Total BST Volume	gallons	2,334,547			
Maximum Storage Capacity (no cone volume considered)					

Average Annual	days	5.1		
Peak 3-Week	days	4.2		
Peak Day	days	3.1		
<u>Dewatering</u>		Flow	TSS	Concentration
Loading		gpm	lbs/day	% solids
Average Annual		294	94,615	2.68%
Max 3-Week		359	115,430	2.68%
Max Week	⇨	448	143,814	2.68%
Peak Day		489	157,060	2.68%
Centridry Dewatering/Drying				
Number of Units	no.	0.0		
Nominal Size	model	CD3074		
Dewatering Capacity	dry lbs/hr	0		
Heat Demand	BTU/hr	0		
Centrifuge Dewatering				
Number of Units Needed (Max Wee	no.	4.0		
Nominal Size	model	CP3074		
Hydraulic Capacity/BFP	gpm	2,500		
Operating Units at Annual Average	no.	1.6		
Operating Units at Max 3-week Con	no.	1.9		
Operating Units at Max Week Cond	no.	2.4		
Operating Units at Peak Day Condi	no.	2.6		
<u>Dewatered Biosolids</u>				
Dry Solids TSS (lbs)				
Average Annual	lbs/day	87,518		
Peak 3-Week	lbs/day	106,772		
Peak Day	lbs/day	145,281		
Dry Solids TSS (tons)				
Average Annual		44		
Peak 3-Week		53		
Peak Week		67		
Peak Day		73		
Dry Solids VSS (lbs)				
Average Annual	lbs/day	49,741		
Peak 3-Week	lbs/day	60,684		
Peak Day	lbs/day	82,570		
Wet Cake from Centridry (lbs)				
Average Annual	wet lbs/day	0		
Peak 3-Week	wet lbs/day	0		
Peak Day	wet lbs/day	0		
Wet Cake from Centrifuges (lbs)				
Average Annual	wet lbs/day	350,074		
Peak 3-Week	wet lbs/day	427,090		
Peak Day	wet lbs/day	581,122		
Wet Cake (lbs)				
Average Annual	wet lbs/day	350,074		
Peak 3-Week	wet lbs/day	427,090		
Peak Day	wet lbs/day	581,122		
Wet Cake (tons)				
Average Annual	wet ton/day	175		
Peak 3-Week	wet ton/day	214		
Peak Day	wet ton/day	291		

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

Design Criteria

Influent Specifications (for year 2010)	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH_3 mg/l
Thickened Solids											
Average Annual	259	373,198	62,500	194,480	49,375	153,639	80,756	251,287	24,735	76,987	681

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Variables

Influent		
Percentage of Total Solids to VERTAD	100.0% %	
Diluted THS Concentration	6.3% %	62,500 mg/l
THS Temperature	15 C	59 F
Dilution Water Temperature	10 C	50 F
Volatile Percentage of Total Solids	79.0% %	
Reactor(s)		
HRT	4 days	
Temperature	60 C	140 F
Oxygen Transfer Efficiency	50.0% %	
Oxygen Requirement	1.4 lbs O_2/lb VS Destroyed	
VS Destruction	40.0% %	
COD Destruction	50.0% %	
FOG Destruction	90.0% %	
Org-N Destruction	45.0% %	
Heat Generation	9000 Btu/lb VS Destroyed	
Biofilter Loading	11.0 m^3/hr-m^2	
Biofilter Temperature	30 C	86 F
Biofilter Off-gas Temperature	30 C	86 F
Product Constituents		
Ammonia (NH_3)	1000 mg/l	3091 lb/day
Ammonium Bicarbonate (NH_4 HCO_3)	4,647 mg/l	
% Bicarbonate Release to Float to this %	25.0% %	
Product Flow per VerTad	89 gpm	
VerTad Internal Solids Concentration	4.4% %	44,132 mg/l
Flotation Thickener		
Float Solids Concentration	6.9% %	68,859 mg/l
Capture Efficiency	95.0% %	
Surface Solids Loading	1.8 lb/ft^2/hr	43.2 lb/ft^2/d
93% Sulfuric Acid Addition	721 mg/l	0.0004 gal H_2SO_4/gal Product
Polymer Split to Flotation Thickeners	0.0% %	
Anaerobic Digester		
HRT	24 days	
Temperature	35 C	95 F
VS Destruction	50.0% %	
COD Destruction	50.0% %	
FOG Destruction	50.0% %	
Org-N Destruction	50.0% %	
Internal TS concentration (Product)	4.5% %	45,000 mg/l
Gas Production	0.54 L CH_4/g COD_{rem}	(Jenny Yoo)
Heat of Combustion of Methane	22,773 Btu/lb CH4	
Specific Volume of Methane	24.2 cf/lb CH4	
Energy Constant	2546 Btu/hr/hp	
Digester Volume	2,731,095 gallons	
Centrifuge		
Cake Solids Concentration	30.0% %	300,000 mg/l
Capture Efficiency	95.0% %	
Polymer Addition	20 lb/dry ton	
Polymer Concentration	0.26% %	2,560 mg/l
Geology		
Ground Temperature	10 C	50 F
Overall Transfer Coeff. (Shaft to Dirt)	0.34 Btu/hr-°F-ft^2	
Overall Transfer Coeff. (Head Tank to A	N/A Btu/hr-°F-ft^3	
Ambient Air		
Air Temperature	15 C	59 F

Equipment Sizing (Average Annual)

Reactor(s)		
THS Flowrate	259 gpm	373,198 gpd

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Dilution Water	0	gpm	0	gpd		
Total Liquid Flow to Reactor(s)	259	gpm	373,198	gpd		
Active Volume Required	199,558	ft^3				
Shaft						
Depth	350	ft				
Diameter	154	in	12.8	ft		
Volume	45,197	ft^3				
Time Requirement at Reactor Temperature	287	min	4.8	hr		
Soak Zone Volume Required	3,513	ft^3				
Soak Zone Depth	27.2	ft				
Soak Zone Safety Factor	1.1					
Actual Soak Zone Depth for Design	29.9	ft				
Actual Soak Zone Volume	3861	ft^3				
Time in Soak Zone	316	min	5.3	hr		
Head Tank						
Sidewater Depth	12.8	ft				
Width	25.6	ft				
Length	76.9	ft				
Head Tank Surface Area	1,973	ft^2				
Active Volume	25,299	ft^3				
Total Active Volume per Reactor	70,496	ft^3				
Number of Reactors Required	2.83					
Compressor(s)						
Percentage of Energy Recoverable from Compress	20%					
TS Loading on the Shaft(s)	194,480	lb/day			3679	kg/hr
Total VS Destroyed	61,456	lb/day			1163	kg/hr
Total Oxygen Requirement	86,038	lb/day	1,036,006	ft^3/day		
OTE	50%					
Total Aeration Requirement	9,866,725	ft^3/day	741,978	lb/day		
Total Aeration Rate	6,852	scfm	1903	hp	1419	kW
Total Aeration Rate per Shaft	2420	scfm				
Compressed Air Temperature	32	C	90	F		
Voidage Check						
Voidage	0.63753	ft^3 per scfm of aeration (at 14 scfm)				
Total Voidage in the Bioreactor(s) plus Head Tank	3,439	ft^3	1.7%	of the active volume		
Shaft Cross-sectional Area	129.1	ft^2				
Riser Cross-sectional Area	96.8	ft^2				
Downcomer Cross-sectional Area	32.3	ft^2				
Riser Liquid Velocity	2.5	ft/s				
Bubble Rise Velocity	1.0	ft/s				
Bulk Riser Velocity	3.5	ft/s				
Riser Flowrate	20,338	ft^3/min				
Aeration at Top of Bioreactor	1,964	scfm				
Voidage at the Top of a Bioreactor	9.7%	%	(Stay below 14%)			
Voidage at the Top of a Bioreactor	18.7	scfm/ft^2	(Stay below 40scfm/ft^2)			
Biofilter(s)						
Total Aeration Rate to Shaft(s)	6,852	scfm				
Biofilter Loading Rate	11.0	m^3/hr-m^2	0.601	ft/min		
Total Biofilter Surface Area Required	11,392	ft^2				
Biofilter Surface Area per Shaft	4,024	ft^2				
Length of Biofilter	76.9	ft				
Width of Biofilter	52.3	ft				
Depth under Media	1	ft				
Media Depth	9	ft				
Standpipe Depth over Media	3	ft				
Active Volume per Biofilter (w/media)	52,314	ft^3				
Biofilter Porosity	40%	%				
Active Liquid Volume per Biofilter	20,926	ft^3				
Total Volume of Biofilter(s)	148,091	ft^3				
Total Condensation in the Biofilter(s)	8.11	gpm				
Condensation per Biofilter	2.87	gpm				
SAFT(s)						
Total Off-Gas Flowrate	1.75	gpm	2,522	gpd		
Total Product Flowrate	257	gpm	370,676	gpd		
Product Concentration	4.4%	%	44,132	mg/l		
Total Sulfuric Flowrate	0.10	gpm	146	gpd	1.11	ton/day
Total Liquid Flow to SAFT(s)	258	gpm	370,822	gpd		
TS Loading to the SAFT(s)	133,024	lb/day	5,543	lb/hr		
Surface Area Required	3,079	ft^2				
Number of SAFTs Required	3					
Surface Area Required per SAFT	1,088	ft^2				
Width	19.0	ft	(Stay below 20ft			
Length	57.1	ft				
Sidewater Depth	12.0	ft				

Total VerTad Digester Volume:
1,492,791 gallons

Goal Seek
Set cell to an even number. Solve for shaft diameter cell above (repeat for desired shaft diameter)

Available energy with 20% recovery:
969,048 Btu/hr

1.2209 kWhr/kg VS destroyed
0.3858 kWhr/kg TS in

Ratio of Volatile to Total Solids in VerTad

6.2%	194,480	lb/day	TSin
1.3%	40840.8	lb/day	FSin
4.9%	153639.2	lb/day	VSin
	0.79		VS/TSin

3.0%	92183.52	lb/day	VSout
4.3%	133024.3	lb/day	TSout
1.3%	40840.8	lb/day	FSout
	0.692982		VS/TSout

Free-board	1.0	ft			
Active Volume per SAFT	13,053	ft^3			
Total Volume of SAFT(s)	40,030	ft^3			
HRT of SAFT(s)	19	hrs			
Product Storage Tank (SAFT Float Solids)					
Total Liquid Flow to SAFT(s)	258	gpm	370,822	gpd	
TS Loading to the SAFT(s)	133,024	lb/day	5,543	lb/hr	
TS Subnatant Return from SAFT(s)	6,651	lb/day	277	lb/hr	
Thickened Solids (TS) from SAFT(s)	126,373	lb/day	5,266	lb/hr	
Thickened Solids from SAFT(s)	153	gpm	220,109	gpd	
Volatile Solids from SAFT(s)	87,574	lb/day	3,649	lb/hr	
Total Subnatant Return from SAFT(s)	105	gpm	150,713	gpd	
Subnatant Return Solids Concentration	5,293	mg/l	0.53%	%	
Underflow Subnatant Return from SAFT(s)	15%	%	(Set by standpipe height in the subnatant trough)		
HRT of Storage Tank(s)	4	hrs			
Total Active Storage Tank Volume Required	4,904	ft^3			
Tank Height	11.0	ft			
Free-board	1.0	ft			
Maximum Sidewater Depth	10.0	ft			
Total Surface Area of Storage Tank(s)	490	ft^2			
Number of Storage Tanks Required	1				
Width of Storage Tank(s)	19	ft			
Length of Storage Tank(s)	25.8	ft			
Anaerobic Digester(s)					
Total Liquid Flow to Digester(s)	153	gpm	220,109	gpd	
Total TS Loading to the Digester(s)	126,373	lb/day	5,266	lb/hr	
Total VS Loading to the Digester(s)	87,574	lb/day	3,649	lb/hr	
Number of Digesters Required	2				
Liquid Flow per Digester	76	gpm	110,054	gpd	
TS Loading per Digester	63,187	lb/day	2,633	lb/hr	
VS Loading per Digester	43,787	lb/day	1,824	lb/hr	
Influent Solids Concentration	6.9%	%	68,859	mg/l	
Desired Internal Solids Concentration	4.5%	%	45,000	mg/l	
Desired Digester HRT	24.0	days			
Actual Digester HRT	24.8	days			
VS out of the Anaerobic Digester	21,894	lb/day	912	lb/hr	
TS out of the Anaerobic Digester	41,293	lb/day	1,721	lb/hr	
Liquid Flow per Digester	76	gpm	110,054	gpd	Goal Seek
Product Solids Concentration	4.5%	%	45,000	mg/l	Set cell equal to value in cell D56,
Total Digester Volume Requirement	5,282,614	gallons			and adjust cell D42
Active Volume per Digester	2,641,307	gallons			
Total Methane Production	15,401,370	L/day CH4			
Total Methane Production	543,897	cf/day CH4			
Total Combined VS Destruction	107,547	lbVS dest/day			
Total Methane Production	5.1	cf CH4/lb VS dest			
Heat Available from Methane	511,814,150	Btu/day			
Heat Available from Methane	21,325,590	Btu/hr			
Methane per Anaerobic Digester	7,700,685	L/day CH4			
Overall Combined VS Destruction	70.0%				
TS Loading from the Anaerobic Digester	82,586	lb/day	3,441	lb/hr	
Anaerobic Product Storage Tank					
Total Liquid Flow to Tank	153	gpm	220,109	gpd	
Anaerobic Product Solids Concentration	4.5%	%			
HRT of Storage Tank(s)	4	hrs			
Total Active Storage Tank Volume Required	4,904	ft^3			
Tank Height	11.0	ft			
Free-board	1.0	ft			
Maximum Sidewater Depth	10.0	ft			
Total Surface Area of Storage Tank(s)	490	ft^2			
Number of Storage Tanks Required	1				
Width of Storage Tank(s)	19	ft			
Length of Storage Tank(s)	25.8	ft			
Centrifuge(s)					
Total Solids from Anaerobic Digester(s)	82,586	lb/day	3,441	lb/hr	
Total Solids from Anaerobic Digester(s)	153	gpm	220,109	gpd	
Polymer Addition	20.0	lb/dry ton			
Made-down Polymer Flowrate	27	gpm	38,691	gpd	
Polymer Flowrate	826	lb/day	34	lb/hr	
Total Flow to Centrifuge(s)	180	gpm	258,800	gpd	
Concentration of Flow to Centrifuge(s)	38,655	mg/l	3.9%	%	
Total Mass Loading on Centrifuge(s)	83,412	lb/day	3,475	lb/hr	
Total Centrate Solids	4,171	lb/day	174	lb/hr	

c.f. methane from straight anaerobic:

34,225,266	L/day CH4
1,208,661	cf/day CH4
92,184	lbVS dest/day
13.1	cf CH4/lb VS dest
1,137,364,779	Btu/day
47,390,199	Btu/hr

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Total Cake Solids	79,241	lb/day	40	ton/day
Total Wet Cake Solids	264,137	Wet lb/day	132	Wet ton/day
Total Cake Flow to Trucks	22.0	gpm	31,679	gpd
Cake Solids Concentration	300,000	mg/l	30%	%
Total Centrate Flow to Plant	158	gpm	227,121	gpd
Centrate Solids Concentration	2,202	mg/l	0.22%	%
Operating units at annual avg loading	1.4			
OTHER				
Centrate Surge Tank Size				
Retention Time Required	10	min		
Centrate Flowrate	158	gpm	227,121	gpd
Total Volume Required	211	ft^3	1,577	gallons
Number of Tanks	2			
Volume Required per Tank	105	ft^3	789	gallons
Tank Height	8	ft		
Tank Diameter	4.1	ft		
Sulfuric Acid Tank Size				
# of Days of Sulfuric Acid Supply Required	7	days		
Total Sulfuric Flowrate	0.10	gpm	146	gpd
Total Volume Required	136	ft^3	1,020	gallons
Tank Height	10	ft		
Tank Diameter	4.2	ft		
Influent Preheat Heat Exchanger				
Heat Recovery per Shaft	1,845,327	Btu/hr		
Heat Transfer Coefficient	75	Btu/hr-°F-ft^2		
Sludge to Sludge Temperature Approach	0	C		
Hot Product Supply	60.0	C	140	F
Cool Influent Sludge	15.0	C	59	F
Tempered Product Return	37.5	C	99.5	F
Preheated Sludge to Shaft	37.5	C	99.5	F
Log Mean Differential Temperature	22.5	C	41	F
# of Heat Exchangers	3			
Surface Area for each Heat Exchanger	607	ft^2		
Flowrate of product sludge through exchanger	91	gpm		
Shaft Internal Recycle Heat Exchanger				
Cooling Requirement per Shaft	4,403,256	Btu/hr		
Heat Transfer Coefficient	75	Btu/hr-°F-ft^2		
Sludge to Water Temperature Approach	5	C		
Sludge Supply	60.0	C	140	F
Cooling Water Supply	10.0	C	50	F
Sludge Return	45.0	C	113	F
Cooling Water Return	50.0	C	122	F
Log Mean Differential Temperature	20.0	C	36	F
# of Heat Exchangers	3			
Surface Area for each Heat Exchanger	1633	ft^2		
Flowrate of reactor sludge through exchanger	245	gpm		
Flowrate of water through exchanger	122	gpm		
Biofilter Heat Exchanger				
Cooling Requirement per Biofilter	1,455,957	Btu/hr		
Biofilter Temperature	30	C		
Heat Transfer Coefficient	75	Btu/hr-°F-ft^2		
Water to Water Temperature Approach	0	C		
Biofilter Supply	30.0	C	86	F
Cooling Water Supply	10.0	C	50	F
Biofilter Return	20.0	C	68	F
Cooling Water Return	20.0	C	68	F
Differential Temperature	10.0	C	18	F
Surface Area for each Heat Exchanger	1077	ft^2		
# of Heat Exchangers	3			
Flowrate of biofilter liquor through exchanger	161	gpm		
Flowrate of water through exchanger	162	gpm		
Turnover time in the biofilter	969	min	16.2	hrs
Energy Recovery and Generation				
Total Heat Recovery from VerTad System	21,809,899	Btu/hr		
Heat Generation from VerTad Internal Recycle	12,464,674	Btu/hr	(As 122°F cooling water return)	
Heat Removal from Biofilter	4,121,501	Btu/hr		
Heat used to Preheat Sludge to VerTad	5,223,724	Btu/hr		
Heat available from the Compressors	969,048	Btu/hr		

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

Design Criteria

Influent Specifications (for year 2019)	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH_3 mg/l
Thickened Solids											
Average Annual	259	373,198	**62,500**	**194,480**	49,375	153,639	**80,756**	251,287	**24,735**	76,987	**681**

Variables

Influent

Percentage of Total Solids to VERTAD	**100.0%**	%		
Diluted THS Concentration	**6.3%**	%	62,500	mg/l
THS Temperature	15	C	59	F
Dilution Water Temperature	10	C	50	F
Volatile Percentage of Total Solids	**79.0%**	%		

Reactor(s)

HRT	**1.37**	days		
Temperature	**60**	C	140	F
Oxygen Transfer Efficiency	**35.0%**	%		
Oxygen Requirement	**1.4**	lbs O_2/lb VS Destroyed		
VS Destruction	**14%**	%		
COD Destruction	**19%**	%		
FOG Destruction	**90%**	%		
Org-N Destruction	**14%**	%		
Heat Generation	9000	Btu/lb VS Destroyed		
Biofilter Loading	**11.0**	m^3/hr-m^2		
Biofilter Temperature	**30**	C	86	F
Biofilter Off-gas Temperature	**30**	C	86	F

Product Constituents

Ammonia (NH_3)	**1000**	mg/l	3101	lb/day
Ammonium Bicarbonate (NH_4 HCO_3)	4,647	mg/l		
% Bicarbonate Release to Float to this %ta	**25.0%**	%		
Product Flow per VerTad	135	gpm		
VerTad Internal Solids Concentration	**5.7%**	%	56,623	mg/l

Flotation Thickener

Float Solids Concentration	**7.3%**	%	72,811	mg/l
Capture Efficiency	**95.0%**	%		
Surface Solids Loading	**1.8**	lb/ft^2/hr	43.2	lb/ft^2/d
93% Sulfuric Acid Addition	721	mg/l	0.0004	gal H_2SO_4/gal Product
Polymer Split to Flotation Thickeners	**0.0%**	%		

Anaerobic Digester

HRT	**24**	days		
Temperature	**35**	C	95	F
VS Destruction	**59.5%**	%		
COD Destruction	**59.5%**	%		
FOG Destruction	**59.5%**	%		
Org-N Destruction	**59.5%**	%		
Internal TS concentration (Product)	**4.5%**	%	45,000	mg/l
Gas Production	**0.54**	L CH_4/g COD_{rem}	(Jenny Yoo)	
Heat of Combustion of Methane	**22,773**	Btu/lb CH4		
Specific Volume of Methane	**24.2**	cf/lb CH4		
Energy Constant	2546	Btu/hr/hp		
Digester Volume	**2,731,095**	gallons		

Centrifuge

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Cake Solids Concentration	**30.0%**	%	300,000	mg/l
Capture Efficiency	**95.0%**	%		
Polymer Addition	20	lb/dry ton		
Polymer Concentration	**0.26%**	%	2,560	mg/l

Geology

Ground Temperature	**10**	C	50	F
Overall Transfer Coeff. (Shaft to Dirt)	**0.34**	Btu/hr-°F-ft^2		
Overall Transfer Coeff. (Head Tank to Air)	**N/A**	Btu/hr-°F-ft^3		

Ambient Air

Air Temperature	**15**	C	59	F

Equipment Sizing (Average Annual)

Reactor(s)

THS Flowrate	259	gpm	373,198	gpd
Dilution Water	0	gpm	0	gpd
Total Liquid Flow to Reactor(s)	259	gpm	373,198	gpd
Active Volume Required	68,115	ft^3		

Shaft

Depth	**350**	ft		
Diameter	**115**	in	9.6	ft
Volume	25,433	ft^3		
Time Requirement at Reactor Temperature	287	min	4.8	hr
Soak Zone Volume Required	5,272	ft^3		
Soak Zone Depth	72.6	ft		
Soak Zone Safety Factor	**1.1**			
Actual Soak Zone Depth for Design	79.8	ft		
Actual Soak Zone Volume	5800	ft^3		
Time in Soak Zone	316	min	5.3	hr

Head Tank

Sidewater Depth	9.6	ft
Width	19.2	ft
Length	57.7	ft
Head Tank Surface Area	1,110	ft^2
Active Volume	10,679	ft^3
Total Active Volume per Reactor	36,113	ft^3
Number of Reactors Required	1.89	

Total VerTad Digester Volume:
509,531 gallons

Goal Seek
Set cell to an even number. Solve for shaft diameter cell above (repeat for desired shaft diameter)

Compressor(s)

Percentage of Energy Recoverable from Compressor	20%					
TS Loading on the Shaft(s)	194,480	lb/day			3679	kg/hr
Total VS Destroyed	20,977	lb/day			397	kg/hr
Total Oxygen Requirement	29,367	lb/day	353,618	ft^3/day		
OTE	35%					
Total Aeration Requirement	4,811,123	ft^3/day	361,796	lb/day		
Total Aeration Rate	3,341	scfm	928	hp	692	kW
Total Aeration Rate per Shaft	1,771	scfm				
Compressed Air Temperature	**32**	C	90	F		

Available energy with 20% recove
472,518 Btu/hr

1.7441 kWhr/kg VS destroyed
0.1881 kWhr/kg TS in

Voidage Check

Voidage	**0.63753**	ft^3 per scfm of aeration (at 14 scfm)	
Total Voidage in the Bioreactor(s) plus Head Tank(s)	2,030	ft^3	3.0% of the active volume
Shaft Cross-sectional Area	72.7	ft^2	
Riser Cross-sectional Area	54.5	ft^2	
Downcomer Cross-sectional Area	18.2	ft^2	
Riser Liquid Velocity	**2.5**	ft/s	
Bubble Rise Velocity	**1.0**	ft/s	
Bulk Riser Velocity	3.5	ft/s	
Riser Flowrate	11,445	ft^3/min	
Aeration at Top of Bioreactor	1,521	scfm	
Voidage at the Top of a Bioreactor	13.3%	%	(Stay below 14%)
Voidage at the Top of a Bioreactor	24.4	scfm/ft^2	(Stay below 40scfm/ft^2)

Biofilter(s)

Total Aeration Rate to Shaft(s)	3,341	scfm		
Biofilter Loading Rate	11.0	$m^3/hr-m^2$	0.601	ft/min
Total Biofilter Surface Area Required	5,555	ft^2		
Biofilter Surface Area per Shaft	2,945	ft^2		
Length of Biofilter	57.7	ft		
Width of Biofilter	51.0	ft		
Depth under Media	**1**	ft		
Media Depth	**9**	ft		
Standpipe Depth over Media	**3**	ft		
Active Volume per Biofilter (w/media)	38,284	ft^3		
Biofilter Porosity	**40%**	%		
Active Liquid Volume per Biofilter	15,314	ft^3		
Total Volume of Biofilter(s)	72,211	ft^3		
Total Condensation in the Biofilter(s)	3.96	gpm		
Condensation per Biofilter	2.10	gpm		

SAFT(s)

Total Off-Gas Flowrate	0.85	gpm	1,230	gpd		
Total Product Flowrate	258	gpm	371,968	gpd		
Product Concentration	5.7%	%	56,623	mg/l		
Total Sulfuric Flowrate	0.10	gpm	146	gpd	1.11	ton/day
Total Liquid Flow to SAFT(s)	258	gpm	372,114	gpd		
TS Loading to the SAFT(s)	173,503	lb/day	7,229	lb/hr		
Surface Area Required	4,016	ft^2				
Number of SAFTs Required	**4**					
Surface Area Required per SAFT	1,004	ft^2				
Width	18.3	ft	(Stay below 20ft)			
Length	54.9	ft				
Sidewater Depth	**12.0**	ft				
Free-board	**1.0**	ft				
Active Volume per SAFT	12,049	ft^3				
Total Volume of SAFT(s)	52,212	ft^3				
HRT of SAFT(s)	25	hrs				

Ratio of Volatile to Total Solids in VerTad

6.2%	194,480	lb/day	TSin
1.3%	40840.8	lb/day	FSin
4.9%	153639	lb/day	VSin
	0.79		VS/TSin
4.3%	132663	lb/day	VSout
5.6%	173503	lb/day	TSout
1.3%	40840.8	lb/day	FSout
	0.76461		VS/TSout

Product Storage Tank (SAFT Float Solids)

Total Liquid Flow to SAFT(s)	258	gpm	372,114	gpd
TS Loading to the SAFT(s)	173,503	lb/day	7,229	lb/hr
TS Subnatant Return from SAFT(s)	8,675	lb/day	361	lb/hr
Thickened Solids (TS) from SAFT(s)	164,828	lb/day	6,868	lb/hr
Thickened Solids from SAFT(s)	189	gpm	271,505	gpd
Volatile Solids from SAFT(s)	126,030	lb/day	5,251	lb/hr
Total Subnatant Return from SAFT(s)	70	gpm	100,609	gpd
Subnatant Return Solids Concentration	10,341	mg/l	1.03%	%
Underflow Subnatant Return from SAFT(s)	**15%**	%	(Set by standpipe height in the subnatant trough)	
HRT of Storage Tank(s)	**4**	hrs		
Total Active Storage Tank Volume Required	6,049	ft^3		
Tank Height	**11.0**	ft		
Free-board	**1.0**	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	605	ft^2		
Number of Storage Tanks Required	**1**			
Width of Storage Tank(s)	18	ft		
Length of Storage Tank(s)	33.1	ft		

Anaerobic Digester(s)

Total Liquid Flow to Digester(s)	189	gpm	271,505	gpd
Total TS Loading to the Digester(s)	164,828	lb/day	6,868	lb/hr
Total VS Loading to the Digester(s)	126,030	lb/day	5,251	lb/hr
Number of Digesters Required	**3**			
Liquid Flow per Digester	63	gpm	90,502	gpd

Active Volume per Digester	2,172,039	gallons		
Total Methane Production	29,817,916	L/day CH4		
Total Methane Production	1,053,016	cf/day CH4		
Total Combined VS Destruction	99,911	lbVS dest/day		
Total Methane Production	10.5	cf CH4/lb VS dest		
Heat Available from Methane	990,900,907	Btu/day		
Heat Available from Methane	41,287,538	Btu/hr		
Methane per Anaerobic Digester	9,939,305	L/day CH4		
Overall Combined VS Destruction	65.0%			
TS Loading from the Anaerobic Digester	89,841	lb/day	3,743	lb/hr

c.f. methane from straight anaerobic:

34,225,266	L/day CH4
1,208,661	cf/day CH4
92,184	lbVS dest/day
13.1	cf CH4/lb VS dest
1,137,364,779	Btu/day
47,390,199	Btu/hr

Anaerobic Product Storage Tank

Total Liquid Flow to Tank	189	gpm	271,505	gpd
Anaerobic Product Solids Concentration	4.0%	%		
HRT of Storage Tank(s)	**4**	hrs		
Total Active Storage Tank Volume Required	6,049	ft^3		
Tank Height	**11.0**	ft		
Free-board	**1.0**	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	605	ft^2		
Number of Storage Tanks Required	**1**			
Width of Storage Tank(s)	18	ft		
Length of Storage Tank(s)	33.1	ft		

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Centrifuge(s)

Total Solids from Anaerobic Digester(s)	89,841	lb/day	3,743	lb/hr
Total Solids from Anaerobic Digester(s)	189	gpm	271,505	gpd
Polymer Addition	20.0	lb/dry ton		
Made-down Polymer Flowrate	29	gpm	42,090	gpd
Polymer Flowrate	898	lb/day	37	lb/hr
Total Flow to Centrifuge(s)	218	gpm	313,595	gpd
Concentration of Flow to Centrifuge(s)	34,703	mg/l	3.5%	%
Total Mass Loading on Centrifuge(s)	90,739	lb/day	3,781	lb/hr
Total Centrate Solids	4,537	lb/day	189	lb/hr
Total Cake Solids	86,202	lb/day	43	ton/day
Total Wet Cake Solids	287,341	Wet lb/day	144	Wet ton/day
Total Cake Flow to Trucks	23.9	gpm	34,462	gpd
Cake Solids Concentration	300,000	mg/l	30%	%
Total Centrate Flow to Plant	194	gpm	279,133	gpd
Centrate Solids Concentration	1,949	mg/l	0.19%	%
Operating units at annual avg loading	1.5			

OTHER

Centrate Surge Tank Size

Retention Time Required	**10**	min		
Centrate Flowrate	194	gpm	279,133	gpd
Total Volume Required	259	ft^3	1,938	gallons
Number of Tanks	**2**			
Volume Required per Tank	130	ft^3	969	gallons
Tank Height	**8**	ft		
Tank Diameter	4.5	ft		

Sulfuric Acid Tank Size

# of Days of Sulfuric Acid Supply Required	**7**	days		
Total Sulfuric Flowrate	0.10	gpm	146	gpd
Total Volume Required	137	ft^3	1,023	gallons
Tank Height	**10**	ft		
Tank Diameter	4.2	ft		

Influent Preheat Heat Exchanger

Heat Recovery per Shaft	2,779,142	Btu/hr		
Heat Transfer Coefficient	**75**	Btu/hr-°F-ft^2		
Sludge to Sludge Temperature Approach	**0**	C		
Hot Product Supply	60.0	C	140	F
Cool Influent Sludge	15.0	C	59	F
Tempered Product Return	37.5	C	99.5	F

Sulfuric Acid Tank Size

# of Days of Sulfuric Acid Supply Required	**7**	days		
Total Sulfuric Flowrate	0.10	gpm	146	gpd
Total Volume Required	137	ft^3	1,023	gallons
Tank Height	**10**	ft		
Tank Diameter	4.2	ft		

Influent Preheat Heat Exchanger

Heat Recovery per Shaft	2,779,142	Btu/hr		
Heat Transfer Coefficient	**75**	Btu/hr-°F-ft^2		
Sludge to Sludge Temperature Approach	**0**	C		
Hot Product Supply	60.0	C	140	F
Cool Influent Sludge	15.0	C	59	F
Tempered Product Return	**37.5**	C	99.5	F
Preheated Sludge to Shaft	37.5	C	99.5	F
Log Mean Differential Temperature	22.5	C	41	F
# of Heat Exchangers	2			
Surface Area for each Heat Exchanger	914	ft^2		
Flowrate of product sludge through exchanger	137	gpm		

Shaft Internal Recycle Heat Exchanger

Cooling Requirement per Shaft	0	Btu/hr		
Heat Transfer Coefficient	**75**	Btu/hr-°F-ft^2		
Sludge to Water Temperature Approach	**5**	C		
Sludge Supply	60.0	C	140	F
Cooling Water Supply	10.0	C	50	F
Sludge Return	**45.0**	C	113	F
Cooling Water Return	50.0	C	122	F
Log Mean Differential Temperature	20.0	C	36	F
# of Heat Exchangers	2			
Surface Area for each Heat Exchanger	0	ft^2		
Flowrate of reactor sludge through exchanger	0	gpm		
Flowrate of water through exchanger	0	gpm		

Biofilter Heat Exchanger

Cooling Requirement per Biofilter	1,065,487	Btu/hr		
Biofilter Temperature	30	C		
Heat Transfer Coefficient	**75**	Btu/hr-°F-ft^2		
Water to Water Temperature Approach	**0**	C		
Biofilter Supply	30.0	C	86	F
Cooling Water Supply	10.0	C	50	F
Biofilter Return	**20.0**	C	68	F
Cooling Water Return	**20.0**	C	68	F
Differential Temperature	10.0	C	18	F
Surface Area for each Heat Exchanger	788	ft^2		
# of Heat Exchangers	2			
Flowrate of biofilter liquor through exchanger	118	gpm		
Flowrate of water through exchanger	118	gpm		
Turnover time in the biofilter	969	min	16.2	hrs

Energy Recovery and Generation

Total Heat Recovery from VerTad System	7,251,621	Btu/hr	
Heat Generation from VerTad Internal Recycle	0	Btu/hr	(As 122°F cooling water return)
Heat Removal from Biofilter	2,009,689	Btu/hr	
Heat used to Preheat Sludge to VerTad	5,241,932	Btu/hr	
Heat available from the Compressors	472,518	Btu/hr	

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

Peaking Factor = **1.52**

Design Criteria

Influent Specifications (for year 2019)	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH$_3$ mg/l
Thickened Solids											
Average Annual	259	373,198	**62,500**	**194,480**	49,375	153,639	**80,756**	251,287	**24,735**	76,987	**681**
Peak Week	418	601,300	**62,500**	**313,348**	49,375	247,545	**80,756**	404,876	**24,735**	124,042	**681**

Variables

Influent

Percentage of Total Solids to VERTAD	**100.0%**	%		
Diluted THS Concentration	**6.25%**	%	62,500	mg/l
THS Temperature	**15**	C	59	F
Dilution Water Temperature	**10**	C	50	F
Volatile Percentage of Total Solids	**79.0%**	%		

Reactor(s)

HRT	**4**	days		
Temperature	**60**	C	140	F
Oxygen Transfer Efficiency	**50.0%**	%		
Oxygen Requirement	**1.4**	lbs O$_2$/lb VS Destroyed		
VS Destruction	**40.0%**	%		
COD Destruction	**50.0%**	%		
FOG Destruction	**90.0%**	%		
Org-N Destruction	**45.0%**	%		
Heat Generation	**9000**	Btu/lb VS Destroyed		
Biofilter Loading	**11.0**	m^3/hr-m^2		
Biofilter Temperature	**30**	C	86	F
Biofilter Off-gas Temperature	**30**	C	86	F

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Product Constituents

Ammonia (NH$_3$)	**1000**	mg/l	4981	lb/day
Ammonium Bicarbonate (NH$_4$ HCO$_3$)	4,647	mg/l		
% Bicarbonate Release to Float to this %tage	**25.0%**	%		
Product Flow per VerTad	81	gpm		
VerTad Internal Solids Concentration	**4.4%**	%	44,068	mg/l

Flotation Thickener

Float Solids Concentration	**10.0%**	%	100,000	mg/l
Capture Efficiency	**95.0%**	%		
Surface Solids Loading	**1.8**	lb/ft^2/hr	43.2	lb/ft^2/d
93% Sulfuric Acid Addition	721	mg/l	0.0004	gal H$_2$SO$_4$/gal Product
Polymer Split to Flotation Thickeners	**0.0%**	%		

Anaerobic Digester

HRT	**24**	days		
Temperature	**35**	C	95	F
VS Destruction	**50.0%**	%		
COD Destruction	**50.0%**	%		
FOG Destruction	**50.0%**	%		
Org-N Destruction	**50.0%**	%		
Internal TS concentration (Product)	**4.5%**	%	45,000	mg/l
Gas Production	**0.4**	L CH$_4$/g COD$_{rem}$	(Jenny Yoo)	
Heat of Combustion of Methane	**22,773**	Btu/lb CH4		
Specific Volume of Methane	**24.2**	cf/lb CH4		
Energy Constant	2546	Btu/hr/hp		
Digester Volume	**2,000,000**	gallons		

Centrifuge

Cake Solids Concentration	**30.0%**	%	300,000	mg/l
Capture Efficiency	**95.0%**	%		
Polymer Addition	**20**	lb/dry ton		
Polymer Concentration	**0.26%**	%	2,560	mg/l

Geology

Ground Temperature	**10** C	50 F
Overall Transfer Coeff. (Shaft to Dirt)	**0.34** Btu/hr-°F-ft²	
Overall Transfer Coeff. (Head Tank to Air)	**N/A** Btu/hr-°F-ft³	

Ambient Air

Air Temperature	**15** C	59 F

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Equipment Sizing (Average Annual)

Reactor(s)

THS Flowrate	418 gpm	601,300 gpd
Dilution Water	0 gpm	0 gpd
Total Liquid Flow to Reactor(s)	418 gpm	601,300 gpd
Active Volume Required	321,529 ft³	
Shaft		
Depth	**350** ft	
Diameter	**148** in	12.3 ft
Volume	41,802 ft³	
Time Requirement at Reactor Temperature	287 min	4.8 hr
Soak Zone Volume Required	3,205 ft³	
Soak Zone Depth	26.8 ft	
Soak Zone Safety Factor	**1.1**	
Actual Soak Zone Depth for Design	29.5 ft	
Actual Soak Zone Volume	3525 ft³	
Time in Soak Zone	316 min	5.3 hr
Head Tank		
Sidewater Depth	12.3 ft	
Width	24.7 ft	
Length	74.0 ft	
Head Tank Surface Area	1,825 ft²	
Active Volume	22,503 ft³	
Total Active Volume per Reactor	64,306 ft³	
Number of Reactors Required	5.00	

Total VerTad Digester Volume:
2,405,199 gallons

Goal Seek
Set cell to an even number, Solve for shaft diameter cell above (repeat for desired shaft diameter)

Compressor(s)

Percentage of Energy Recoverable from Compressor	20%		
TS Loading on the Shaft(s)	313,348 lb/day		5927 kg/hr
Total VS Destroyed	99,018 lb/day		1873 kg/hr
Total Oxygen Requirement	138,625 lb/day	1,669,223 ft³/day	
OTE	50%		
Total Aeration Requirement	15,897,361 ft³/day	1,195,482 lb/day	
Total Aeration Rate	11,040 scfm	3067 hp	2287 kW
Total Aeration Rate per Shaft	2208 scfm		
Compressed Air Temperature	**32** C	90 F	

Available energy with 20% recovery:
1,561,339 Btu/hr

1.22086 kWhr/kg VS destroyed
0.38579 kWhr/kg TS in

Voidage Check

Voidage	**0.63753** ft³ per scfm of aeration (at 14 scfm)	
Total Voidage in the Bioreactor(s) plus Head Tank(s)	4,250 ft³	1.3% of the active volume
Shaft Cross-sectional Area	119.4 ft²	
Riser Cross-sectional Area	89.6 ft²	
Downcomer Cross-sectional Area	29.9 ft²	
Riser Liquid Velocity	**2.5** ft/s	
Bubble Rise Velocity	**1.0** ft/s	
Bulk Riser Velocity	3.5 ft/s	
Riser Flowrate	18,811 ft³/min	
Aeration at Top of Bioreactor	1,808 scfm	
Voidage at the Top of a Bioreactor	9.6% %	(Stay below 14%)
Voidage at the Top of a Bioreactor	18.5 scfm/ft²	(Stay below 40scfm/ft²)

Biofilter(s)

Total Aeration Rate to Shaft(s)	11,040 scfm	
Biofilter Loading Rate	11.0 m³/hr-m²	0.601 ft/min
Total Biofilter Surface Area Required	18,354 ft²	
Biofilter Surface Area per Shaft	3,671 ft²	
Length of Biofilter	74.0 ft	
Width of Biofilter	49.6 ft	
Depth under Media	**1** ft	
Media Depth	**9** ft	
Standpipe Depth over Media	**3** ft	
Active Volume per Biofilter (w/media)	47,721 ft³	

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Biofilter Porosity	**40%**	%		
Active Liquid Volume per Biofilter	19,088	ft^{3}		
Total Volume of Biofilter(s)	238,605	ft^{3}		
Total Condensation in the Biofilter(s)	12.49	gpm		
Condensation per Biofilter	2.50	gpm		

SAFT(s)

Total Off-Gas Flowrate	2.70	gpm	3,881	gpd	
Total Product Flowrate	415	gpm	597,418	gpd	
Product Concentration	4.4%	%	44,068	mg/l	
Total Sulfuric Flowrate	0.16	gpm	235	gpd	1.79 ton/day
Total Liquid Flow to SAFT(s)	415	gpm	597,653	gpd	
TS Loading to the SAFT(s)	214,330	lb/day	8,930	lb/hr	
Surface Area Required	4,961	ft^{2}			
Number of SAFTs Required	5				
Surface Area Required per SAFT	992	ft^{2}			
Width	18.2	ft	(Stay below 20ft)		
Length	54.6	ft			
Sidewater Depth	**12.0**	ft			
Free-board	**1.0**	ft			
Active Volume per SAFT	11,907	ft^{3}			
Total Volume of SAFT(s)	64,497	ft^{3}			
HRT of SAFT(s)	19	hrs			

Ratio of Volatile to Total Solids in VerTad

6.2%	313,348	lb/day	TSin
1.3%	65803.08	lb/day	FSin
4.9%	247544.9	lb/day	VSin
	0.79		VS/TSin
3.0%	148527	lb/day	VSout
4.3%	214330	lb/day	TSout
1.3%	65803.08	lb/day	FSout
	0.692982		VS/TSout

Product Storage Tank (SAFT Float Solids)

Total Liquid Flow to SAFT(s)	415	gpm	597,653	gpd
TS Loading to the SAFT(s)	214,330	lb/day	8,930	lb/hr
TS Subnatant Return from SAFT(s)	10,717	lb/day	447	lb/hr
Thickened Solids (TS) from SAFT(s)	203,614	lb/day	8,484	lb/hr
Thickened Solids from SAFT(s)	170	gpm	244,203	gpd
Volatile Solids from SAFT(s)	141,101	lb/day	5,879	lb/hr
Total Subnatant Return from SAFT(s)	245	gpm	353,450	gpd
Subnatant Return Solids Concentration	3,636	mg/l	0.36%	%
Underflow Subnatant Return from SAFT(s)	**15%**	%	(Set by standpipe height in the subnatant trough)	
HRT of Storage Tank(s)	**4**	hrs		
Total Active Storage Tank Volume Required	5,441	ft^{3}		
Tank Height	**11.0**	ft		
Free-board	**1.0**	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	544	ft^{2}		
Number of Storage Tanks Required	**1**			
Width of Storage Tank(s)	18	ft		
Length of Storage Tank(s)	29.9	ft		

Centrifuge(s)

Total Solids from VerTad	203,614	lb/day	8,484	lb/hr
Total Solids from VerTad	170	gpm	244,203	gpd
Polymer Addition	20.0	lb/dry ton		
Made-down Polymer Flowrate	66	gpm	95,392	gpd
Polymer Flowrate	2,036	lb/day	85	lb/hr
Total Flow to Centrifuge(s)	236	gpm	339,595	gpd
Concentration of Flow to Centrifuge(s)	72,629	mg/l	7.3%	%
Total Mass Loading on Centrifuge(s)	205,650	lb/day	8,569	lb/hr
Total Centrate Solids	10,282	lb/day	428	lb/hr
Total Cake Solids	195,367	lb/day	98	ton/day
Total Wet Cake Solids	651,224	Wet lb/day	326	Wet ton/day
Total Cake Flow to Trucks	54.2	gpm	78,104	gpd
Cake Solids Concentration	300,000	mg/l	30%	%
Total Centrate Flow to Plant	182	gpm	261,490	gpd
Centrate Solids Concentration	4,716	mg/l	0.47%	%
Operating units at annual avg loading	3.4			

OTHER

Centrate Surge Tank Size

Retention Time Required	**10**	min		
Centrate Flowrate	182	gpm	261,490	gpd
Total Volume Required	243	ft^{3}	1,816	gallons
Number of Tanks	2			

Volume Required per Tank	121 ft^3	908 gallons
Tank Height	**8** ft	
Tank Diameter	4.4 ft	

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Sulfuric Acid Tank Size

# of Days of Sulfuric Acid Supply Required	**7** days	
Total Sulfuric Flowrate	0.16 gpm	235 gpd
Total Volume Required	220 ft^3	1,644 gallons
Tank Height	**10** ft	
Tank Diameter	5.3 ft	

Influent Preheat Heat Exchanger

Heat Recovery per Shaft	1,683,815 Btu/hr	
Heat Transfer Coefficient	**75** Btu/hr-°F-ft^2	
Sludge to Sludge Temperature Approach	**0** C	
Hot Product Supply	60.0 C	140 F
Cool Influent Sludge	15.0 C	59 F
Tempered Product Return	**37.5** C	99.5 F
Preheated Sludge to Shaft	37.5 C	99.5 F
Log Mean Differential Temperature	22.5 C	41 F
# of Heat Exchangers	5	
Surface Area for each Heat Exchanger	554 ft^2	
Flowrate of product sludge through exchanger	83 gpm	

Shaft Internal Recycle Heat Exchanger

Cooling Requirement per Shaft	4,072,531 Btu/hr	
Heat Transfer Coefficient	**75** Btu/hr-°F-ft^2	
Sludge to Water Temperature Approach	**5** C	
Sludge Supply	60.0 C	140 F
Cooling Water Supply	10.0 C	50 F
Sludge Return	**45.0** C	113 F
Cooling Water Return	50.0 C	122 F
Log Mean Differential Temperature	20.0 C	36 F
# of Heat Exchangers	5	
Surface Area for each Heat Exchanger	1510 ft^2	
Flowrate of reactor sludge through exchanger	226 gpm	
Flowrate of water through exchanger	113 gpm	

Biofilter Heat Exchanger

Cooling Requirement per Biofilter	1,344,933 Btu/hr	
Biofilter Temperature	30 C	
Heat Transfer Coefficient	**75** Btu/hr-°F-ft^2	
Water to Water Temperature Approach	**0** C	
Biofilter Supply	30.0 C	86 F
Cooling Water Supply	10.0 C	50 F
Biofilter Return	**20.0** C	68 F
Cooling Water Return	**20.0** C	68 F
Differential Temperature	10.0 C	18 F
Surface Area for each Heat Exchanger	995 ft^2	
# of Heat Exchangers	5	
Flowrate of biofilter liquor through exchanger	149 gpm	
Flowrate of water through exchanger	149 gpm	
Turnover time in the biofilter	957 min	16.0 hrs

Energy Recovery and Generation

Total Heat Recovery from VerTad System	35,506,373 Btu/hr	
Heat Generation from VerTad Internal Recycle	20,362,643 Btu/hr	(As 122°F cooling water return)
Heat Removal from Biofilter	6,724,659 Btu/hr	
Heat used to Preheat Sludge to VerTad	8,419,071 Btu/hr	
Heat available from the Compressors	1,561,339 Btu/hr	

Table 4- : Energy Cost Analysis - Digester Gas and Hot Water

Heating value of gas	BTU/cft	600
Net value of digester gas	$/therm	$ 0.07
Net value of hot water	$/therm	$ 0.02
Average boiler efficiency	percent	70%

Average Annual Values	Units	Alternative 2 Digester 5	Alternative 3 TPAD	Alternative 6 TPAD+Class A	Alt SP1 4 day / 2 D	Alt SP2 1.4 day / 3 D	Alt SP3 4 day VERTAD
Number of Anaerobic Digesters		5	4	4	2	3	0
Digester Gas Production	cft/day	1,302,092	1,497,982	1,497,982	543,897	1,053,016	0
Heating Value @ 600 BTU/cft	BTU/day	781,255,320	898,789,320	898,789,320	326,338,411	631,809,470	-
Heating Value	Therms/yr	2,851,582	3,280,581	3,280,581	1,191,135	2,306,105	-
Heat Required for Digestion	BTU/hr	0	12,080,252	12,480,252	6,363,914	6,654,013	0
Hot water from VERTAD	BTU/hr	-	-	-	(21,809,899)	(7,251,621)	(23,359,456)
Total Heat Required	BTU/hr	-	12,080,252	12,480,252	(15,445,985)	(597,608)	(23,359,456)
Excess Heat Available	BTU/hr	-			15,445,985	597,608	23,359,456
	Therms/yr	-			1,353,068	52,350	2,046,288
Gas Demand	BTU/hr	-	17,257,502	17,828,931	-	-	-
Gas Demand	Therms/yr	-	1,511,757	1,561,814	-	-	-
Excess Gas Available for Sale	Therms/yr	2,851,582	1,768,824	1,718,767	1,191,135	2,306,105	-
Annual Revenue from Gas Sold	$/yr	$ 199,611	$ 123,818	$ 120,314	$ 83,379	$ 161,427	$ -
Annual value of excess hot water	$/yr	$ -	$ -	$ -	$ 27,061	$ 1,047	$ 40,926

- Anaerobic Digester Heat Demand is generated by the heat extractors and accounted for as part of electrical demand

Table 4- : CHEMICAL COST ANALYSIS

Process Area	Alternative 2 Digester 5 4 Centrifuges	Alternative 3 Thermo-meso 4 Centrifuges	Alternative 6 Class A 4 Centrifuges	Alt SP1 4 day / 2 D 4 Centrifuges	Alt SP2 1.4 day / 3 D 4 Centrifuges	Alt SP3 4 day VERTAD 4 Centrifuges
VERTAD flotation	0	0	0	$ 40,521	$ 40,662	$ 42,966
Centrifuge Dewatering	$ 1,136,043	998,257	998,257	$ 678,237	$ 737,817	$ 880,093
Total annual cost	$ 1,136,043	$ 998,257	$ 998,257	$ 718,758	$ 778,479	$ 923,058

Polymer usage for centrifuge is 25% greater than BFPs				Polymer dose
Acid Cost is	$	100.00	/ ton	lb/dt
Polymer cost is	$	1.80	/ lb polymer	
anaerobic	$	62.50	/dt of biosolids	35
Vertad-anaer.	$	45.00	/dt of biosolids	25
Vertad	$	36.00	/dt of biosolids	20

Table 4: Operation and Maintenance Costs - Equipment Maintenance

Equipment Name	Equip Purchase Cost	Annual Mainten. percent	Annual Mainten. Cost	Operating Equipment Alt 2 Dig 5	Alt 3 Thermo-meso	Alt 6 Class A	Alt SP1 4 d/ 2 Dig	Alt SP2 1.4 d/ 3 Dig	Alt SP3 4 day VERT	Annual Maintenance Cost Alt 2 Digester 5	Alt 3 Thermo-meso	Alt 6 Class A	Alt SP1 4 d / 2 Dig	Alt SP2 1.4 d/ 3 Dig	Alt SP3 4 day VERT
Blending Tank Equipment	($)	(%)	$/yr	No.	No.	No.	No.	No.	No.	$/yr	$/yr	$/yr	$/yr	$/yr	$/yr
Blending tank circ pump	35,000	5%	1,750	2	1	1	1	1	1	3,500	1,750	1,750	1,750	1,750	1,750
Digester feed pump	25,000	10%	2,500	5	2	2	2	3	0	12,500	5,000	5,000	5,000	7,500	-
Digester withdrawal pumps	25,000	10%	2,500	6	6	7	2	3	0	15,000	15,000	17,500	5,000	7,500	-
Digester Equipment															
Digester mixing compresso	87,000	5%	4,350	5	4	4	2	3	0	21,750	17,400	17,400	8,700	13,050	-
Grinders	15,000	10%	1,500	10	9	9	4	6	0	15,000	13,500	13,500	6,000	9,000	-
Circ sludge pump 1	20,000	5%	1,000	5	3	3	2	3	0	5,000	3,000	3,000	2,000	3,000	-
Circ sludge pump 2 (hex)	20,000	5%	1,000	5	6	6	2	3	0	5,000	6,000	6,000	2,000	3,000	-
Heating System Equipment															
Heat extractors	350,000	5%	17,500	2	0	0	0	0	0	35,000	-	-	-	-	-
HWRS pump	20,000	5%	1,000	4	7	7	2	2	0	4,000	7,000	7,000	2,000	2,000	-
Boilers	60,000	10%	6,000	0	2	2	0	0	0	-	12,000	12,000	-	-	-
Heat exchangers	30,000	3%	900		3	4					2,700	3,600			
Hold Tank Equipment															
Grinder	15,000	10%	1,500	0	0	1	0	0	0	-	-	1,500	-	-	-
Circ sludge pump (hex	25,000	5%	1,250	0	0	1	0	0	0	-	-	1,250	-	-	-
HWRS pump	20,000	5%	1,000	0	0	1	0	0	0	-	-	1,000	-	-	-
Tank Withdrawal Pum	25,000	10%	2,500	0	0	3	0	0	0	-	-	7,500	-	-	-
Automatic Valves	32,000	10%	3,200	0	0	1	0	0	0	-	-	3,200	-	-	-
VERTAD Equipment															
Supply pumps	25,000	10%	2,500	0	0	0	3	2	5	-	-	-	7,500	5,000	12,500
Anaerobic feed pumps	25,000	10%	2,500	0	0	0	2	3	0	-	-	-	5,000	7,500	-
Product Tank Mixers	5,000	10%	500	0	0	0	1	3	1	-	-	-	500	1,500	500
Compressors - 1040 hp	150,000	5%	7,500	0	0	0	2	0	0	-	-	-	15,000	-	-
208 hp	35,000	5%	1,750	0	0	0	0	2	0	-	-	-	-	3,500	-
641 hp	120,000	5%	6,000	0	0	0	0	0	5	-	-	-	-	-	30,000
Boiler pumps	150,000	2%	3,000	0	0	0	3	2	5	-	-	-	9,000	6,000	15,000
Flotation Tank Equipment															
Acid pumps	15,000	10%	1,500	0	0	0	1	1	1	-	-	-	1,500	1,500	1,500
Subnatant pump	25,000	5%	1,250	0	0	0	2	1	5	-	-	-	2,500	1,250	6,250
Scraper	20,000	5%	1,000	0	0	0	6	8	1	-	-	-	6,000	8,000	1,000
Heat Exchangers	25,000	10%	2,500	0	0	0	9	4	15	-	-	-	22,500	10,000	37,500
Subtotal (Digestion)										116,750	83,350	101,200	101,950	91,050	106,000
Dewatering Equipment															
Centrifuge	500,000	6%	30,000	2	2	2	2	2	4	60,000	60,000	60,000	60,000	60,000	120,000
Centrifuge feed pump	45,000	10%	4,500	2	2	2	2	2	4	9,000	9,000	9,000	9,000	9,000	18,000
Polymer feed pump	7,500	10%	750	2	2	2	2	2	4	1,500	1,500	1,500	1,500	1,500	3,000
Centridry	1,000,000	5%	50,000	0	0	0	0	0	0	-	-	-	-	-	-
Subtotal (Dewatering)										70,500	70,500	70,500	70,500	70,500	141,000
Total Annual Maintenance Cost										$ 187,250	$ 153,850	$ 171,700	$ 172,450	$ 161,550	$ 247,000

Table 4- : Operation and maintenance Costs - Labor Costs

Table 4: Operation and Maintenance Costs - Labor Costs

Operating Labor for Digestion								
Digester Labor (1995)	$ 282,793							
Inflate to 1999 dollars	$ 295,776							
Divide by 5.5 tanks	$ 53,777							
(Strategy is that each digester is a tank, the BST is a tank, each VERTAD reactor is 3/4 tank, each flotation tank is 1/2 tank and each Blending tank is 1/2 a tank)								
Annual Labor Cost			Alt 1	Alt 2	Alt 3	Alt 6	Alt SP1	Alt SP3
		No of tanks	5.5	7	5.5	7.5	7.25	7.75
		Annual Labor	$ 295,776	$ 376,442	$ 295,776	$ 403,331	$ 389,886	$ 416,775
Operating Labor for Centrifuge Dewatering								
Estimated Operating Cost per		No. of Centrifuges's	2	2	2	2	2	4
Centrifuge installed	$ 90,000	Annual Labor	$ 180,000	$ 180,000	$ 180,000	$ 180,000	$ 180,000	$ 360,000

Table 4- : Electrical Costs

	HP		Operating Equipment						Constant Electricity Usage						Variable Electricity Usage					
Equipment Name	(name plate)	draw (est)	Alt 2 Dig 5 No.	Alt 3 Thermo-meso No.	Alt 6 Class A No.	Alt SP1 4 d/ 2 Dig No.	Alt SP2 1.4 d/ 3 Dig No.	Alt SP3 4 day VERT Class A	Alt 2 Dig 5 HP draw	Alt 3 Thermo-meso HP draw	Alt 6 Class A HP draw	Alt SP1 4 d/ 2 Dig HP draw	Alt SP2 1.4 d/ 3 Dig HP draw	Alt SP3 4 day VERT HP draw	Alt 2 Dig 5 HP draw	Alt 3 Thermo-meso HP draw	Alt 6 Class A HP draw	Alt SP1 4 d/ 2 Dig HP draw	Alt SP2 1.4 d/ 3 Dig HP draw	Alt SP3 4 day VERT HP draw
Blending Tank Equipment																				
Blending Tank Circ pump	30	27	2	1	1	1	1	1	54	27	27	27	27	27						
Digester Feed pump	40	38	5	5	5	2	3	0							190	190	190	76	114	0
Digester Withdrawal Pumps		37	5	6	8	2	3	0							185	222	296	74	111	0
Digester Equipment (number digesters)			5	4	4	2	3	0												
Digester Mixing Compressors	60	57	5	4	4	2	3	0	285	228	228	114	171	0						
Grinders	10	8	10	9	9	4	12	12	80	72	72	32	96	96						
Circ sludge pump 1	60	57	5	3	3	2	6	6	285	171	171	114	342	342						
Circ sludge pump 2 (hex)	50	47	5	6	6	2	6	6	235	282	282	94	282	282						
Heating System Equipment																				
Heat extractors																				
Hp/MMBTUH	68	68																		
MMBTUH required			7.6	0	0	0	0	0												
HP draw															517	0	0	0	0	0
HWRS pump	30	27	5	6	6	2	2	0	135	162	162	54	54	0						
Hold Tank Equipment																				
Grinders	10	8	0	0	1						8									
Circ sludge pump (hex)	50	40	0	0	1						40									
HWRS pump	30	20	0	0	1						20									
Tank Withdrawal Pumps	25	15	0	0	3						45									
VERTAD Equipment (Reactors)						3	2	5												
Supply Pumps	5	3.8				3	2	5	0			11.4	7.6	19						
Anaerobic feed pumps	3	2.8				2	3	0	0						0			5.6	8.4	0
Product Tank Mixer	10	9				1	3	1	0			9	27	9						
Compressors	1040	1030				2			0						0			2060	0	0
	208	200					2												400	
	641	412						5												2,060
Biofilter pumps	6	5.7				3	2	5	0			17.1	11.4	28.5						
Flotation Equipment																				
Acid Pump	0.5	0.5				1	1	1				0.5	0.5	0.5						
Subnatant Pump	2	2				2	1	5				4	2	10						
Scraper	1	1				6	8	1				6	8	1						
Heat exchangers	0	0				9	4	3				0	0	0						
Total Hp Digestion									1,074	942	1,055	483	1,029	815	892	412	486	2,216	633	2,060
Total electricity draw (kW)									801	702	787	360	767	608	665	307	362	1,652	472	1,536
Total cost/yr (Digestion)									$ 319,917	$ 280,598	$ 314,257	$ 143,873	$ 308,364	$ 242,768	$ 265,644	$ 122,724	$ 144,767	$ 659,970	$ 188,674	$ 613,621

Dewatering Equipment																				
Centrifuges main drive	125	100	1.8	1.6	1.6	1.4	1.5	3.4							179	158	158	140	150	340
Centrifuges back drive	30	18	1.8	1.6	1.6	1.4	1.5	3.4							32	28	28	25	27	61
Centridry	638	477	0.0	0.0	0.0	0.0	0.0	0.0							0	0	0	0	0	0
Centrifuge feed sludge pump	30	20	1.8	1.6	1.6	1.4	1.5	3.4							36	32	32	28	30	68
Polymer feed pump	2	1	1.8	1.6	1.6	1.4	1.5	3.4							2	2	2	1	2	3
Conveyors	5	3	2.0	2.0	2.0	2.0	2.0	2.0	6	6	6	6	6	6						
Total Hp Dewatering									6	6	6	6	6	6	249	219	220	195	209	473
Total electricity draw (kW)									4	4	4	4	4	4	186	163	164	145	155	352
Total cost/yr Dewatering									$ 1,787	$ 1,787	$ 1,787	$ 1,787	$ 1,787	$ 1,787	$ 74,303	$ 65,291	$ 65,532	$ 57,966	$ 62,107	$ 140,775

Cost of Electricity	0.0456 $/kW-hr

CAPITAL AND ANNUAL COST ANALYSIS

Capital Cost Estimate

Digestion

Category		TPAD	Class A	Alt SP1	Alt SP2	Alt SP3
Site Work		140,000	140,000	606,000	606,000	606,000
Demolition		-	-			
Structures		1,820,000	3,300,000	1,330,000	890,000	1,940,000
Equipment & Mech		2,270,000	3,310,000	1,060,000	720,000	1,400,000
Patent Fee		-	-	-	-	-
Electrical/ I&C		460,000	660,000	320,000	270,000	390,000
Testing/ Start-up		110,000	120,000	incl	incl	incl
NORAM provided equipment and engineering				6,290,000	3,900,000	8,470,000
Cased Reactors				6,140,000	3,140,000	7,000,000
Subtotal		4,800,000	7,530,000	15,746,000	9,526,000	19,806,000
Contractor Indirects, OH(	35%	1,680,000	2,635,500	-	-	-
NORAM estimate	11.6%			1,096,896	652,616	1,314,976
NORAM provided	5%			314,500	195,000	423,500
Subtotal		6,480,000	10,165,500	17,157,396	10,373,616	21,544,476
Contingency	30%	1,944,000	3,049,650	1,545,341	1,089,072	1,982,299
B&C Digester 5	20%					
Drilling	15%			1,003,554	512,909	1,142,164
NORAM provided	10%			685,381	424,702	921,346
Subtotal		8,424,000	13,215,150	20,391,672	12,400,299	25,590,285
Sales tax	8.4%	707,616	1,110,073	1,712,900	1,041,625	2,149,584
Subtotal		9,131,616	14,325,223	22,104,572	13,441,924	27,739,869
Allied Cost (35%)	35%	3,196,066	5,013,828	7,736,600	4,704,673	9,708,954
Less NORAM engineerin	15%			1,160,490	705,701	1,456,343
Total		12,327,682	19,339,051	28,680,682	17,440,896	35,992,480

Dewatering

Category		Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Site Work		20,000	20,000	20,000	20,000	20,000
Demolition						
Structures		780,000	780,000	780,000	780,000	1,170,000
Equipment & Mech		2,950,000	2,950,000	2,950,000	2,950,000	4,425,000
Patent Fee						
Electrical/ I&C		750,000	750,000	750,000	750,000	1,125,000
Testing/ Start-up		100,000	100,000	100,000	100,000	100,000
Subtotal		4,600,000	4,600,000	4,600,000	4,600,000	6,840,000
Contractor Indirects, OH(	23%	1,053,400	1,053,400	1,053,400	1,053,400	1,566,360
Subtotal		5,653,400	5,653,400	5,653,400	5,653,400	8,406,360
Contingency	13%	729,289	729,289	729,289	729,289	1,084,420
Subtotal		6,382,689	6,382,689	6,382,689	6,382,689	9,490,780
Sales tax	8.4%	536,146	536,146	536,146	536,146	797,226
Subtotal		6,918,834	6,918,834	6,918,834	6,918,834	10,288,006
Allied Cost (35%)	35%	2,421,592	2,421,592	2,421,592	2,421,592	3,600,802
Total		9,340,426	9,340,426	9,340,426	9,340,426	13,888,808

0.6725146

Total Capital Cost

	Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Grand Total Capital Expenditure	$ 21,668,108	$ 28,679,477	$ 38,021,109	$ 26,781,323	$ 49,881,288

Annual Costs at Design Year Loading, 2010 (todays dollars)

Digestion

Category	97 actuals	Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Equipment Maintenance	155,000	$ 83,350	$ 101,200	$ 101,950	$ 91,050	$ 106,000
Operations Labor	146,000	$ 295,776	$ 403,331	$ 389,886	$ 376,442	$ 416,775
Power						
Fixed	185000	$ 280,598	$ 314,257	$ 143,873	$ 306,364	$ 242,768
Variable		$ 122,724	$ 144,767	$ 659,970	$ 188,674	$ 613,621
Chemicals (acid)		$ -	$ -	$ 40,521	$ 40,662	$ 42,966
Hot water avoided cost		$ -	$ -	$ (27,061)	$ (1,047)	$ (40,926)
Gas Sale Net Revenue		$ (123,818)	$ (120,314)	$ (83,379)	$ (161,427)	$ -
Total Annual	$ 486,000	$ 658,630	$ 843,241	$ 1,225,760	$ 840,717	$ 1,381,204

Dewatering

Category		Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Equipment Maintenance	480000	$ 70,500	$ 70,500	$ 70,500	$ 70,500	$ 141,000
Operations Labor	182000	$ 360,000	$ 360,000	$ 360,000	$ 360,000	$ 540,000
Power						
Fixed	69000	$ 1,787	$ 1,787	$ 1,787	$ 1,787	$ 1,787
Variable		$ 65,291	$ 65,532	$ 57,966	$ 62,107	$ 140,775
Chemicals (polymer)	642000	$ 998,257	$ 998,257	$ 678,237	$ 737,817	$ 880,093
Total Annual	$ 1,373,000	$ 1,495,836	$ 1,496,077	$ 1,168,491	$ 1,232,211	$ 1,703,655

Biosolids Haul and Application

Category	Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Biosolids Haul and Application	$ 2,147,291	$ 2,147,291	$ 1,620,172	$ 1,762,497	$ 3,994,494
Wet Tons	$ 63,888	$ 63,888	$ 48,205	$ 52,440	$ 78,190
$ / WT	$ 33.61	$ 33.61	$ 33.61	$ 33.61	$ 33.61
Dry Tons	15,972	15,972	14,462	15,732	35,655
$ / DT	$ 134.44	$ 134.44	$ 112.03	$ 112.03	$ 112.03

Total Annual Cost (Year 2019, Todays Dollars)

Category	Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Total Annual Cost	$ 4,301,757	$ 4,486,609	$ 4,014,423	$ 3,835,425	$ 7,079,353
Annual Costs expected to not vary with flow	$ 1,092,011	$ 1,251,075	$ 1,067,997	$ 1,206,143	$ 1,448,330
Annual Costs expected to vary with flow	$ 3,209,746	$ 3,235,534	$ 2,946,426	$ 2,629,282	$ 5,631,023

Annual Costs by Year *

Year	Annual Average Flow	Percentage of Year 2010 flow	Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
2003	94.2	90.6%	4,000,040	4,182,469	3,737,459	3,588,273	6,550,037
2004	95.6	91.9%	4,041,767	4,224,531	3,775,763	3,622,453	6,623,240
2005	97.0	93.3%	4,086,704	4,269,828	3,817,013	3,659,263	6,702,074
2006	98.4	94.6%	4,128,430	4,311,890	3,855,316	3,693,444	6,775,278
2007	99.8	96.0%	4,173,367	4,357,188	3,896,566	3,730,254	6,854,112
2008	101.2	97.3%	4,215,093	4,399,250	3,934,870	3,764,435	6,927,315
2009	102.6	98.7%	4,260,030	4,444,547	3,976,120	3,801,245	7,006,149
2010	104.0	100.0%	4,301,757	4,486,609	4,014,423	3,835,425	7,079,353
2011	95.0	92.2%	4,051,396	4,234,237	3,784,602	3,630,341	6,640,133
2012	97.0	93.1%	4,080,284	4,263,357	3,811,120	3,654,005	6,690,812
2013	99.0	93.9%	4,105,962	4,289,241	3,834,691	3,675,039	6,735,860
2014	101.0	94.8%	4,134,850	4,318,361	3,861,209	3,698,703	6,786,540
2015	102.5	95.7%	4,163,737	4,347,481	3,887,727	3,722,366	6,837,219
2016	104.0	96.5%	4,189,415	4,373,365	3,911,298	3,743,400	6,882,267
2017	105.5	97.4%	4,218,303	4,402,485	3,937,816	3,767,064	6,932,946
2018	107.0	98.3%	4,247,191	4,431,605	3,964,334	3,790,728	6,983,625
2019	108.5	99.1%	4,272,869	4,457,489	3,987,905	3,811,762	7,028,674

* Annual cost for each year is calculated as the sum of the portion of the annual cost not expected to vary with flow and the prorated annual cost expected to vary with flow.

Avoided Costs

	Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Digester					
Year in which a new digester is requ	2024	2024	2040	2030	2050
Years away	25	25	41	31	51
Inflated cost	$. 22,103,517	$ 22,103,517	$ 35,469,656	$ 26,392,755	$ 47,668,251
Present Worth	$ 5,150,089	$ 5,150,089	$ 3,253,243	$ 4,335,138	$ 2,441,350
Off-set Capital Cost	$ (1,146,353)	$ (1,146,353)	$ (3,043,199)	$ (1,961,305)	$ (3,855,092)

Capital cost of a digester in 1999 dollars
Capital cost for 2 belt filter presses including
Bldg expansion 1999 dollars

Table 4- : PRESENT WORTH COST ANALYSIS -Life Cycle Costs

		Alt 2	Alt 3	Alt 6	Alt SP1	Alt SP2	Alt SP3
Inflated Annual Costs	inflation						
Year	years						
2003	4	$ 4,757,083	$ 4,502,081	$ 4,707,406	$ 4,206,543	$ 4,038,633	$ 7,372,124
2004	5	$ 4,952,380	$ 4,685,516	$ 4,897,389	$ 4,377,144	$ 4,199,416	$ 7,678,150
2005	6	$ 5,159,279	$ 4,879,738	$ 5,098,398	$ 4,557,713	$ 4,369,352	$ 8,002,627
2006	7	$ 5,369,844	$ 5,077,448	$ 5,303,081	$ 4,741,553	$ 4,542,470	$ 8,332,737
2007	8	$ 5,592,819	$ 5,286,696	$ 5,519,555	$ 4,936,053	$ 4,725,374	$ 8,682,584
2008	9	$ 5,819,788	$ 5,499,741	$ 5,740,023	$ 5,134,112	$ 4,911,733	$ 9,038,575
2009	10	$ 6,060,030	$ 5,725,124	$ 5,973,100	$ 5,343,572	$ 5,108,555	$ 9,415,679
2010	11	$ 6,304,619	$ 5,954,637	$ 6,210,516	$ 5,556,900	$ 5,309,126	$ 9,799,480
2011	12	$ 6,105,727	$ 5,776,322	$ 6,037,010	$ 5,395,937	$ 5,175,999	$ 9,467,242
2012	13	$ 6,335,014	$ 5,992,035	$ 6,260,884	$ 5,596,758	$ 5,366,029	$ 9,825,683
2013	14	$ 6,567,286	$ 6,210,636	$ 6,487,863	$ 5,800,314	$ 5,558,826	$ 10,188,593
2014	15	$ 6,813,230	$ 6,441,961	$ 6,727,866	$ 6,015,638	$ 5,762,458	$ 10,573,208
2015	16	$ 7,068,019	$ 6,681,576	$ 6,976,431	$ 6,238,660	$ 5,973,305	$ 10,971,729
2016	17	$ 7,326,196	$ 6,924,465	$ 7,228,507	$ 6,464,780	$ 6,187,271	$ 11,375,339
2017	18	$ 7,599,443	$ 7,181,379	$ 7,494,936	$ 6,703,868	$ 6,413,174	$ 11,802,877
2018	19	$ 7,882,491	$ 7,447,475	$ 7,770,846	$ 6,951,483	$ 6,647,064	$ 12,245,829
2019	20	$ 8,169,381	$ 7,717,276	$ 8,050,721	$ 7,202,601	$ 6,884,466	$ 12,694,566
PW Annual Costs		$ 50,382,959	$ 47,632,491	$ 49,739,987	$ 44,477,372	$ 42,594,674	$ 78,199,590
PW Capital Costs		$ 17,947,883	$ 21,668,108	$ 28,679,477	$ 38,021,109	$ 26,781,323	$ 49,881,288
Subtotal		$ 68,330,842	$ 69,300,599	$ 78,419,464	$ 82,498,481	$ 69,375,996	$ 128,080,877
PW of Avoided Capital Costs		$ (1,961,305)	$ (1,146,353)	$ (1,146,353)	$ (3,043,199)	$ (1,961,305)	$ (3,855,092)
Total Present Worth		$ 66,370,000	$ 68,154,000	$ 77,273,000	$ 79,455,000	$ 67,415,000	$ 124,226,000

APPENDIX D (c)

West Point Treatment Plant Cost Estimates

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

Design Criteria

Influent Specifications **(for year ??)**	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH_3 mg/l
Thickened Solids											
Average Annual	333	479,738	55,000	220,000	44,000	176,000	80,756	323,024	24,735	98,965	681

Variables

Influent

Percentage of Total Solids to VERTAD	100.0%	%		
Diluted THS Concentration	5.5%	%	55,000	mg/l
THS Temperature	15	C	59	F
Dilution Water Temperature	10	C	50	F
Volatile Percentage of Total Solids	80.0%	%		

Reactor(s)

HRT	4	days		
Temperature	60	C	140	F
Oxygen Transfer Efficiency	50.0%	%		
Oxygen Requirement	1.4	lbs O_2/lb VS Destroyed		
VS Destruction	40.0%	%		
COD Destruction	50.0%	%		
FOG Destruction	90.0%	%		
Org-N Destruction	45.0%	%		
Heat Generation	9000	Btu/lb VS Destroyed		
Biofilter Loading	11.0	$m^3/hr\text{-}m^2$		
Biofilter Temperature	30	C	86	F
Biofilter Off-gas Temperature	30	C	86	F

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Product Constituents

Ammonia (NH_3)	1000	mg/l	3977	lb/day
Ammonium Bicarbonate (NH_4 HCO_3)	4,647	mg/l		
% Bicarbonate Release to Float to this %tage	25.0%	%		
Product Flow per VerTad	81	gpm		
VerTad Internal Solids Concentration	3.8%	%	38,411	mg/l

Flotation Thickener

Float Solids Concentration	7.0%	%	69,545	mg/l
Capture Efficiency	95.0%	%		
Surface Solids Loading	1.8	$lb/ft^2/hr$	43.2	$lb/ft^2/d$
93% Sulfuric Acid Addition	721	mg/l	0.0004	gal H_2SO_4/gal Product
Polymer Split to Flotation Thickeners	0.0%	%		

Anaerobic Digester

HRT	24	days		
Temperature	35	C	95	F
VS Destruction	50.0%	%		
COD Destruction	50.0%	%		
FOG Destruction	50.0%	%		
Org-N Destruction	50.0%	%		
Internal TS concentration (Product)	4.5%	%	45,000	mg/l
Gas Production	0.54	L CH_4/g COD_{rem}	(Jenny Yoo)	
Heat of Combustion of Methane	22,773	Btu/lb CH4		
Specific Volume of Methane	24.2	cf/lb CH4		
Energy Constant	2546	Btu/hr/hp		
Digester Volume	2,000,000	gallons		

Centrifuge

Cake Solids Concentration	30.0%	%	300,000	mg/l
Capture Efficiency	95.0%	%		
Polymer Addition	20	lb/dry ton		
Polymer Concentration	0.26%	%	2,560	mg/l

Geology

Ground Temperature	10	C	50	F
Overall Transfer Coeff. (Shaft to Dirt)	0.34	Btu/hr-°F-ft^2		
Overall Transfer Coeff. (Head Tank to Air)	N/A	Btu/hr-°F-ft^3		

Ambient Air

Air Temperature	15	C	59	F

Equipment Sizing (Average Annual)

Reactor(s)

THS Flowrate	333	gpm	479,738	gpd

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Dilution Water	0	gpm	0	gpd			
Total Liquid Flow to Reactor(s)	333	gpm	479,738	gpd			
Active Volume Required	256,527	ft^3					
Shaft							
Depth	350	ft					
Diameter	148	in	12.3	ft			
Volume	41,703	ft^3					
Time Requirement at Reactor Temperature	287	min	4.8	hr			
Soak Zone Volume Required	3,196	ft^3					
Soak Zone Depth	26.8	ft					
Soak Zone Safety Factor	1.1						
Actual Soak Zone Depth for Design	29.5	ft					
Actual Soak Zone Volume	3515	ft^3					
Time in Soak Zone	316	min	5.3	hr			
Head Tank							
Sidewater Depth	12.3	ft					
Width	24.6	ft					
Length	73.9	ft					
Head Tank Surface Area	1,820	ft^2					
Active Volume	22,423	ft^3					
Total Active Volume per Reactor	64,125	ft^3					
Number of Reactors Required	4.00						
Compressor(s)							
Percentage of Energy Recoverable from Compressor	20%						
TS Loading on the Shaft(s)	220,000	lb/day			4162	kg/hr	
Total VS Destroyed	70,400	lb/day			1332	kg/hr	
Total Oxygen Requirement	98,560	lb/day	1,186,788	ft^3/day			
OTE	50%						
Total Aeration Requirement	11,302,738	ft^3/day	849,966	lb/day			
Total Aeration Rate	7,849	scfm	2180	hp	1626	kW	
Total Aeration Rate per Shaft	1962	scfm					
Compressed Air Temperature	32	C	90	F			
Voidage Check							
Voidage	0.63753	ft^3 per scfm of aeration (at 14 scfm)					
Total Voidage in the Bioreactor(s) plus Head Tank(s)	3,718	ft^3	1.4%	of the active volume			
Shaft Cross-sectional Area	119.2	ft^2					
Riser Cross-sectional Area	89.4	ft^2					
Downcomer Cross-sectional Area	29.8	ft^2					
Riser Liquid Velocity	2.5	ft/s					
Bubble Rise Velocity	1.0	ft/s					
Bulk Riser Velocity	3.5	ft/s					
Riser Flowrate	18,766	ft^3/min					
Aeration at Top of Bioreactor	1,607	scfm					
Voidage at the Top of a Bioreactor	8.6%	%					(Stay below 14%)
Voidage at the Top of a Bioreactor	16.5	scfm/ft^2					(Stay below 40scfm/ft^2)
Biofilter(s)							
Total Aeration Rate to Shaft(s)	7,849	scfm					
Biofilter Loading Rate	11.0	m^3/hr-m^2	0.601	ft/min			
Total Biofilter Surface Area Required	13,050	ft^2					
Biofilter Surface Area per Shaft	3,262	ft^2					
Length of Biofilter	73.9	ft					
Width of Biofilter	44.1	ft					
Depth under Media	1	ft					
Media Depth	9	ft					
Standpipe Depth over Media	3	ft					
Active Volume per Biofilter (w/media)	42,407	ft^3					
Biofilter Porosity	40%	%					
Active Liquid Volume per Biofilter	16,963	ft^3					
Total Volume of Biofilter(s)	169,644	ft^3					
Total Condensation in the Biofilter(s)	8.77	gpm					
Condensation per Biofilter	2.19	gpm					
SAFT(s)							
Total Off-Gas Flowrate	1.89	gpm	2,722	gpd			
Total Product Flowrate	331	gpm	477,016	gpd			
Product Concentration	3.8%	%	38,411	mg/l			
Total Sulfuric Flowrate	0.13	gpm	187	gpd	1.43	ton/day	
Total Liquid Flow to SAFT(s)	331	gpm	477,204	gpd			
TS Loading to the SAFT(s)	149,600	lb/day	6,233	lb/hr			
Surface Area Required	3,463	ft^2					
Number of SAFTs Required	4						
Surface Area Required per SAFT	866	ft^2					
Width	17.0	ft					(Stay below 20ft)
Length	51.0	ft					
Sidewater Depth	12.0	ft					

Total VerTad Digester Volume:
1,918,952 gallons

Goal Seek
Set cell to an even number. Solve for shaft diameter cell above (repeat for desired shaft diameter)

Available energy with 20% recovery:
1,110,084 Btu/hr

1.220859 kWhr/kg VS destroyed
0.390675 kWhr/kg TS in

Ratio of Volatile to Total Solids in VerTad

5.5%	220,000	lb/day	TSin
1.1%	44000	lb/day	FSin
4.4%	176000	lb/day	VSin
	0.8		VS/TSin

Free-board	1.0	ft		
Active Volume per SAFT	10,388	ft^3		
Total Volume of SAFT(s)	45,019	ft^3		
HRT of SAFT(s)	17	hrs		

2.6%	105600	lb/day	VSout
3.7%	149600	lb/day	TSout
1.1%	44000	lb/day	FSout
	0.7058824		VS/TSout

Product Storage Tank (SAFT Float Solids)

Total Liquid Flow to SAFT(s)	331	gpm	477,204	gpd
TS Loading to the SAFT(s)	149,600	lb/day	6,233	lb/hr
TS Subnatant Return from SAFT(s)	7,480	lb/day	312	lb/hr
Thickened Solids (TS) from SAFT(s)	142,120	lb/day	5,922	lb/hr
Thickened Solids from SAFT(s)	170	gpm	245,093	gpd
Volatile Solids from SAFT(s)	100,320	lb/day	4,180	lb/hr
Total Subnatant Return from SAFT(s)	161	gpm	232,111	gpd
Subnatant Return Solids Concentration	3,865	mg/l	0.39%	%
Underflow Subnatant Return from SAFT(s)	15%	%	(Set by standpipe height in the subnatant trough)	
HRT of Storage Tank(s)	4	hrs		
Total Active Storage Tank Volume Required	5,461	ft^3		
Tank Height	11.0	ft		
Free-board	1.0	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	546	ft^2		
Number of Storage Tanks Required	1			
Width of Storage Tank(s)	17	ft		
Length of Storage Tank(s)	32.1	ft		

Anaerobic Digester(s)

Total Liquid Flow to Digester(s)	170	gpm	245,093	gpd
Total TS Loading to the Digester(s)	142,120	lb/day	5,922	lb/hr
Total VS Loading to the Digester(s)	100,320	lb/day	4,180	lb/hr
Number of Digesters Required	3			
Liquid Flow per Digester	57	gpm	81,698	gpd
TS Loading per Digester	47,373	lb/day	1,974	lb/hr
VS Loading per Digester	33,440	lb/day	1,393	lb/hr
Influent Solids Concentration	7.0%	%	69,545	mg/l
Desired Internal Solids Concentration	4.5%	%	45,000	mg/l
Desired Digester HRT	24.0	days		
Actual Digester HRT	24.5	days		
VS out of the Anaerobic Digester	16,720	lb/day	697	lb/hr
TS out of the Anaerobic Digester	30,653	lb/day	1,277	lb/hr
Liquid Flow per Digester	57	gpm	81,698	gpd
Product Solids Concentration	4.5%	%	45,000	mg/l
Total Digester Volume Requirement	5,882,227	gallons		
Active Volume per Digester	1,960,742	gallons		
Total Methane Production	19,798,141	L/day CH4		
Total Methane Production	699,169	cf/day CH4		
Total Combined VS Destruction	123,200	lbVS dest/day		
Total Methane Production	5.7	cf CH4/lb VS dest		
Heat Available from Methane	657,926,458	Btu/day		
Heat Available from Methane	27,413,602	Btu/hr		
Methane per Anaerobic Digester	6,599,380	L/day CH4		
Overall Combined VS Destruction	70.0%			
TS Loading from the Anaerobic Digester	91,960	lb/day	3,832	lb/hr

Goal Seek
Set cell equal to value in cell E196,
and adjust cell D61

Goal Seek
Set cell equal to value in cell D56,
and adjust cell D42

c.f. methane from straight anaerobic:

43,995,869	L/day CH4
1,553,708	cf/day CH4
105,600	lbVS dest/day
14.7	cf CH4/lb VS dest
1,462,058,796	Btu/day
60,919,117	Btu/hr

Anaerobic Product Storage Tank

Total Liquid Flow to Tank	170	gpm	245,093	gpd
Anaerobic Product Solids Concentration	4.5%	%		
HRT of Storage Tank(s)	4	hrs		
Total Active Storage Tank Volume Required	5,461	ft^3		
Tank Height	11.0	ft		
Free-board	1.0	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	546	ft^2		
Number of Storage Tanks Required	1			
Width of Storage Tank(s)	17	ft		
Length of Storage Tank(s)	32.1	ft		

Centrifuge(s)

Total Solids from Anaerobic Digester(s)	91,960	lb/day	3,832	lb/hr
Total Solids from Anaerobic Digester(s)	170	gpm	245,093	gpd
Polymer Addition	20.0	lb/dry ton		
Made-down Polymer Flowrate	30	gpm	43,083	gpd
Polymer Flowrate	920	lb/day	38	lb/hr
Total Flow to Centrifuge(s)	200	gpm	288,175	gpd
Concentration of Flow to Centrifuge(s)	38,655	mg/l	3.9%	%
Total Mass Loading on Centrifuge(s)	92,880	lb/day	3,870	lb/hr
Total Centrate Solids	4,644	lb/day	193	lb/hr

Total Cake Solids	88,236	lb/day	44	ton/day
Total Wet Cake Solids	294,119	Wet lb/day	147	Wet ton/day
Total Cake Flow to Trucks	24.5	gpm	35,275	gpd
Cake Solids Concentration	300,000	mg/l	30%	%
Total Centrate Flow to Plant	176	gpm	252,901	gpd
Centrate Solids Concentration	2,202	mg/l	0.22%	%

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Centrate Surge Tank Size

Retention Time Required	10	min		
Centrate Flowrate	176	gpm	252,901	gpd
Total Volume Required	235	ft^3	1,756	gallons
Number of Tanks	2			
Volume Required per Tank	117	ft^3	878	gallons
Tank Height	8	ft		
Tank Diameter	4.3	ft		

Sulfuric Acid Tank Size

# of Days of Sulfuric Acid Supply Required	7	days		
Total Sulfuric Flowrate	0.13	gpm	187	gpd
Total Volume Required	175	ft^3	1,312	gallons
Tank Height	10	ft		
Tank Diameter	4.7	ft		

Influent Preheat Heat Exchanger

Heat Recovery per Shaft	1,680,409	Btu/hr		
Heat Transfer Coefficient	75	Btu/hr-°F-ft^2		
Sludge to Sludge Temperature Approach	0	C		
Hot Product Supply	60.0	C	140	F
Cool Influent Sludge	15.0	C	59	F
Tempered Product Return	37.5	C	99.5	F
Preheated Sludge to Shaft	37.5	C	99.5	F
Log Mean Differential Temperature	22.5	C	41	F
# of Heat Exchangers	4			
Surface Area for each Heat Exchanger	553	ft^2		
Flowrate of product sludge through exchanger	83	gpm		

Shaft Internal Recycle Heat Exchanger

Cooling Requirement per Shaft	3,320,875	Btu/hr		
Heat Transfer Coefficient	75	Btu/hr-°F-ft^2		
Sludge to Water Temperature Approach	5	C		
Sludge Supply	60.0	C	140	F
Cooling Water Supply	10.0	C	50	F
Sludge Return	45.0	C	113	F
Cooling Water Return	50.0	C	122	F
Log Mean Differential Temperature	20.0	C	36	F
# of Heat Exchangers	4			
Surface Area for each Heat Exchanger	1231	ft^2		
Flowrate of reactor sludge through exchanger	185	gpm		
Flowrate of water through exchanger	92	gpm		

Biofilter Heat Exchanger

Cooling Requirement per Biofilter	1,180,218	Btu/hr		
Biofilter Temperature	30	C		
Heat Transfer Coefficient	75	Btu/hr-°F-ft^2		
Water to Water Temperature Approach	0	C		
Biofilter Supply	30.0	C	86	F
Cooling Water Supply	10.0	C	50	F
Biofilter Return	20.0	C	68	F
Cooling Water Return	20.0	C	68	F
Differential Temperature	10.0	C	18	F
Surface Area for each Heat Exchanger	873	ft^2		
# of Heat Exchangers	4			
Flowrate of biofilter liquor through exchanger	131	gpm		
Flowrate of water through exchanger	131	gpm		
Turnover time in the biofilter	969	min	16.2	hrs

Energy Recovery and Generation

Total Heat Recovery from VerTad System	24,728,503	Btu/hr	
Heat Generation from VerTad Internal Recycle	13,284,840	Btu/hr	(As 122°F cooling water return)
Heat Removal from Biofilter	4,721,348	Btu/hr	
Heat used to Preheat Sludge to VerTad	6,722,315	Btu/hr	
Heat available from the Compressors	1,110,084	Btu/hr	(As 185°F cooling water return)

COLOR CODING LEGEND:
CONSTANTS
INPUTS, VARIABLES
RESULTS, CALCULATED CELLS
GOAL SEEK VALUE

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Design Criteria

Influent Specifications (for year ??)

	Flow gpm	Flow gpd	TS mg/l	TS lbs/day	VS mg/l	VS lbs/day	COD mg/l	COD lbs/day	FOG mg/kg	FOG lbs/day	NH_3 mg/l
Thickened Solids											
Average Annual	333	479,738	55,000	220,000	44,000	176,000	80,756	323,024	24,735	98,965	681

Variables

Influent

Percentage of Total Solids to VERTAD	100.0% %	
Diluted THS Concentration	5.5% %	55,000 mg/l
THS Temperature	15 C	59 F
Dilution Water Temperature	10 C	50 F
Volatile Percentage of Total Solids	80.0% %	

Reactor(s)

HRT	1.5 days	
Temperature	60 C	140 F
Oxygen Transfer Efficiency	40.0% %	
Oxygen Requirement	1.4 lbs O_2/lb VS Destroyed	
VS Destruction	15% %	1
COD Destruction	20% %	
FOG Destruction	90% %	
Org-N Destruction	15% %	
Heat Generation	9000 Btu/lb VS Destroyed	
Biofilter Loading	11.0 m^3/hr-m^2	
Biofilter Temperature	30 C	86 F
Biofilter Off-gas Temperature	30 C	86 F

Product Constituents

Ammonia (NH_3)	1000 mg/l	3990 lb/day
Ammonium Bicarbonate (NH_4 HCO_3)	4,647 mg/l	
% Bicarbonate Release to Float to this %tage	25.0% %	
Product Flow per VerTad	165 gpm	
VerTad Internal Solids Concentration	4.9% %	49,177 mg/l

Flotation Thickener

Float Solids Concentration	7.3% %	73,232 mg/l
Capture Efficiency	95.0% %	
Surface Solids Loading	1.8 lb/ft^2/hr	43.2 lb/ft^2/d
93% Sulfuric Acid Addition	721 mg/l	0.0004 gal H_2SO_4/gal Product
Polymer Split to Flotation Thickeners	0.0% %	

Anaerobic Digester

HRT	24 days	
Temperature	35 C	95 F
VS Destruction	58.5% %	
COD Destruction	58.5% %	
FOG Destruction	58.5% %	
Org-N Destruction	58.5% %	
Internal TS concentration (Product)	4.5% %	45,000 mg/l
Gas Production	0.54 L CH_4/g COD	(Jenny Yoo)
Heat of Combustion of Methane	22,773 Btu/lb CH4	
Specific Volume of Methane	24.2 cf/lb CH4	
Energy Constant	2546 Btu/hr/hp	
Digester Volume	2,000,000 gallons	

Centrifuge

Cake Solids Concentration	30.0% %	300,000 mg/l
Capture Efficiency	95.0% %	
Polymer Addition	20 lb/dry ton	
Polymer Concentration	0.26% %	2,560 mg/l

Geology

Ground Temperature	10 C	50 F
Overall Transfer Coeff. (Shaft to Dirt)	0.34 Btu/hr-°F-ft^2	
Overall Transfer Coeff. (Head Tank to Air)	N/A Btu/hr-°F-ft^3	

Ambient Air

Air Temperature	15 C	59 F

Equipment Sizing (Average Annual)

Reactor(s)

THS Flowrate	333 gpm	479,738 gpd

Dilution Water	0	gpm	0	gpd		
Total Liquid Flow to Reactor(s)	333	gpm	479,738	gpd		
Active Volume Required	93,471	ft^3				
Shaft						
Depth	350	ft				
Diameter	129	in	10.8	ft		
Volume	31,809	ft^3				
Time Requirement at Reactor Temperature	287	min	4.8	hr		
Soak Zone Volume Required	6,393	ft^3				
Soak Zone Depth	70.3	ft				
Soak Zone Safety Factor	1.1					
Actual Soak Zone Depth for Design	77.4	ft				
Actual Soak Zone Volume	7033	ft^3				
Time in Soak Zone	316	min	5.3	hr		
Head Tank						
Sidewater Depth	10.8	ft				
Width	21.5	ft				
Length	64.5	ft				
Head Tank Surface Area	1,389	ft^2				
Active Volume	14,937	ft^3				
Total Active Volume per Reactor	46,747	ft^3				
Number of Reactors Required	2.00					

NB: Could use a single 15ft diameter reactor

Total VerTad Digester Volume:
699,210 gallons

Goal Seek
Set cell to an even number. Solve for shaft diameter cell above (repeat for desired shaft diameter)

Compressor(s)						
Percentage of Energy Recoverable from Compressor	20%					
TS Loading on the Shaft(s)	220,000	lb/day			4162	kg/hr
Total VS Destroyed	25,652	lb/day			485	kg/hr
Total Oxygen Requirement	35,912	lb/day	432,431	ft^3/day		
OTE	40%					
Total Aeration Requirement	5,147,987	ft^3/day	387,129	lb/day		
Total Aeration Rate	3,575	scfm	993	hp	741	kW
Total Aeration Rate per Shaft	1,788	scfm				
Compressed Air Temperature	32	C	90	F		
Voidage Check						
Voidage	0.63753	ft^3 per scfm of aeration (at 14 scfm)				
Total Voidage in the Bioreactor(s) plus Head Tank(s)	2,145	ft^3	2.3%	of the active volume		
Shaft Cross-sectional Area	90.9	ft^2				
Riser Cross-sectional Area	68.2	ft^2				
Downcomer Cross-sectional Area	22.7	ft^2				
Riser Liquid Velocity	2.5	ft/s				
Bubble Rise Velocity	1.0	ft/s				
Bulk Riser Velocity	3.5	ft/s				
Riser Flowrate	14,314	ft^3/min				
Aeration at Top of Bioreactor	1,505	scfm				
Voidage at the Top of a Bioreactor	10.5%	%	(Stay below 14%)			
Voidage at the Top of a Bioreactor	19.7	scfm/ft^2	(Stay below 40scfm/ft^2)			

Available energy with 20% recovery:
505,603 Btu/hr

1.526074 kWhr/kg VS destroyed
0.177938 kWhr/kg TS in

Biofilter(s)						
Total Aeration Rate to Shaft(s)	3,575	scfm				
Biofilter Loading Rate	11.0	m^3/hr-m^2	0.601	ft/min		
Total Biofilter Surface Area Required	5,944	ft^2				
Biofilter Surface Area per Shaft	2,973	ft^2				
Length of Biofilter	64.5	ft				
Width of Biofilter	46.1	ft				
Depth under Media	1	ft				
Media Depth	9	ft				
Standpipe Depth over Media	3	ft				
Active Volume per Biofilter (w/media)	38,643	ft^3				
Biofilter Porosity	40%	%				
Active Liquid Volume per Biofilter	15,457	ft^3				
Total Volume of Biofilter(s)	77,267	ft^3				
Total Condensation in the Biofilter(s)	3.99	gpm				
Condensation per Biofilter	2.00	gpm				
SAFT(s)						
Total Off-Gas Flowrate	0.86	gpm	1,241	gpd		
Total Product Flowrate	332	gpm	478,497	gpd		
Product Concentration	4.9%	%	49,177	mg/l		
Total Sulfuric Flowrate	0.13	gpm	188	gpd	1.43	ton/day
Total Liquid Flow to SAFT(s)	332	gpm	478,685	gpd		
TS Loading to the SAFT(s)	194,348	lb/day	8,098	lb/hr		
Surface Area Required	4,499	ft^2				
Number of SAFTs Required	4					
Surface Area Required per SAFT	1,125	ft^2				
Width	19.4	ft	(Stay below 20ft)			
Length	58.1	ft				
Sidewater Depth	10.0	ft				
Free-board	1.0	ft				

Ratio of Volatile to Total Solids in VerTad

5.5%	220,000	lb/day	TSin
1.1%	44000	lb/day	FSin
4.4%	176000	lb/day	VSin
	0.8		VS/TSin
3.8%	150348.29	lb/day	VSout

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Active Volume per SAFT	11,247	ft^3		
Total Volume of SAFT(s)	49,487	ft^3		
HRT of SAFT(s)	19	hrs		
Product Storage Tank (SAFT Float Solids)				
Total Liquid Flow to SAFT(s)	332	gpm	478,685	gpd
TS Loading to the SAFT(s)	194,348	lb/day	8,098	lb/hr
TS Subnatant Return from SAFT(s)	9,717	lb/day	405	lb/hr
Thickened Solids (TS) from SAFT(s)	184,631	lb/day	7,693	lb/hr
Thickened Solids from SAFT(s)	210	gpm	302,376	gpd
Volatile Solids from SAFT(s)	142,831	lb/day	5,951	lb/hr
Total Subnatant Return from SAFT(s)	122	gpm	176,309	gpd
Subnatant Return Solids Concentration	6,610	mg/l	0.66%	%
Underflow Subnatant Return from SAFT(s)	15%	%	(Set by standpipe height in the subnatant trough)	
HRT of Storage Tank(s)	4	hrs		
Total Active Storage Tank Volume Required	6,737	ft^3		
Tank Height	11.0	ft		
Free-board	1.0	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	674	ft^2		
Number of Storage Tanks Required	1			
Width of Storage Tank(s)	19	ft		
Length of Storage Tank(s)	34.8	ft		
Anaerobic Digester(s)				
Total Liquid Flow to Digester(s)	210	gpm	302,376	gpd
Total TS Loading to the Digester(s)	184,631	lb/day	7,693	lb/hr
Total VS Loading to the Digester(s)	142,831	lb/day	5,951	lb/hr
Number of Digesters Required	4			
Liquid Flow per Digester	52	gpm	75,594	gpd
TS Loading per Digester	46,158	lb/day	1,923	lb/hr
VS Loading per Digester	35,708	lb/day	1,488	lb/hr
Influent Solids Concentration	7.3%	%	73,232	mg/l
Desired Internal Solids Concentration	4.5%	%	45,000	mg/l
Desired Digester HRT	24.0	days		
Actual Digester HRT	26.5	days		
VS out of the Anaerobic Digester	14,819	lb/day	617	lb/hr
TS out of the Anaerobic Digester	25,269	lb/day	1,053	lb/hr
Liquid Flow per Digester	52	gpm	75,594	gpd
Product Solids Concentration	4.0%	%	40,090	mg/l
Total Digester Volume Requirement	7,257,029	gallons		
Active Volume per Digester	1,814,257	gallons		
Total Methane Production	37,259,088	L/day CH4		
Total Methane Production	1,315,800	cf/day CH4		
Total Combined VS Destruction	113,605	lbVS dest/day		
Total Methane Production	11.6	cf CH4/lb VS dest		
Heat Available from Methane	1,238,183,909	Btu/day		
Heat Available from Methane	51,590,996	Btu/hr		
Methane per Anaerobic Digester	9,314,772	L/day CH4		
Overall Combined VS Destruction	64.5%			
TS Loading from the Anaerobic Digester	101,075	lb/day	4,211	lb/hr
Anaerobic Product Storage Tank				
Total Liquid Flow to Tank	210	gpm	302,376	gpd
Anaerobic Product Solids Concentration	4.0%	%		
HRT of Storage Tank(s)	4	hrs		
Total Active Storage Tank Volume Required	6,737	ft^3		
Tank Height	11.0	ft		
Free-board	1.0	ft		
Maximum Sidewater Depth	10.0	ft		
Total Surface Area of Storage Tank(s)	674	ft^2		
Number of Storage Tanks Required	1			
Width of Storage Tank(s)	19	ft		
Length of Storage Tank(s)	34.8	ft		
Centrifuge(s)				
Total Solids from Anaerobic Digester(s)	101,075	lb/day	4,211	lb/hr
Total Solids from Anaerobic Digester(s)	210	gpm	302,376	gpd
Polymer Addition	20.0	lb/dry ton		
Made-down Polymer Flowrate	33	gpm	47,353	gpd
Polymer Flowrate	1,011	lb/day	42	lb/hr
Total Flow to Centrifuge(s)	243	gpm	349,729	gpd
Concentration of Flow to Centrifuge(s)	35,009	mg/l	3.5%	%
Total Mass Loading on Centrifuge(s)	102,086	lb/day	4,254	lb/hr
Total Centrate Solids	5,104	lb/day	213	lb/hr
Total Cake Solids	96,981	lb/day	48	ton/day

4.9%	194348.29	lb/day	TSout
1.1%	44000	lb/day	FSout
	0.7736023		VS/TSout

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Goal Seek
Set cell equal to value in cell E196.
and adjust cell D61

Goal Seek
Set cell equal to value in cell D56.
and adjust cell D42

c.f. methane from straight anaerobic:

43,995,869	L/day CH4
1,553,708	cf/day CH4
105,600	lbVS dest/day
14.7	cf CH4/lb VS dest
1,462,058,796	Btu/day
60,919,117	Btu/hr

Total Wet Cake Solids	323,271 Wet lb/day	162 Wet ton/day
Total Cake Flow to Trucks	26.9 gpm	38,771 gpd
Cake Solids Concentration	300,000 mg/l	30% %
Total Centrate Flow to Plant	216 gpm	310,958 gpd
Centrate Solids Concentration	1,969 mg/l	0.20% %

OTHER

Centrate Surge Tank Size

Retention Time Required	10 min	
Centrate Flowrate	216 gpm	310,958 gpd
Total Volume Required	289 ft^3	2,159 gallons
Number of Tanks	1	
Volume Required per Tank	289 ft^3	2,159 gallons
Tank Height	8 ft	
Tank Diameter	6.8 ft	

Sulfuric Acid Tank Size

# of Days of Sulfuric Acid Supply Required	7 days	
Total Sulfuric Flowrate	0.13 gpm	188 gpd
Total Volume Required	176 ft^3	1,317 gallons
Tank Height	10 ft	
Tank Diameter	4.7 ft	

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Influent Preheat Heat Exchanger

Heat Recovery per Shaft	3,372,395 Btu/hr	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Sludge to Sludge Temperature Approach	0 C	
Hot Product Supply	60.0 C	140 F
Cool Influent Sludge	15.0 C	59 F
Tempered Product Return	37.5 C	99.5 F
Preheated Sludge to Shaft	37.5 C	99.5 F
Log Mean Differential Temperature	22.5 C	41 F
# of Heat Exchangers	2	
Surface Area for each Heat Exchanger	1109 ft^2	
Flowrate of product sludge through exchanger	166 gpm	

Shaft Internal Recycle Heat Exchanger

Cooling Requirement per Shaft	0 Btu/hr	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Sludge to Water Temperature Approach	5 C	
Sludge Supply	60.0 C	140 F
Cooling Water Supply	10.0 C	50 F
Sludge Return	45.0 C	113 F
Cooling Water Return	50.0 C	122 F
Log Mean Differential Temperature	20.0 C	36 F
# of Heat Exchangers	2	
Surface Area for each Heat Exchanger	0 ft^2	
Flowrate of reactor sludge through exchanger	0 gpm	
Flowrate of water through exchanger	0 gpm	

Biofilter Heat Exchanger

Cooling Requirement per Biofilter	1,075,459 Btu/hr	
Biofilter Temperature	30 C	
Heat Transfer Coefficient	75 Btu/hr-°F-ft^2	
Water to Water Temperature Approach	0 C	
Biofilter Supply	30.0 C	86 F
Cooling Water Supply	10.0 C	50 F
Biofilter Return	**20.0** C	68 F
Cooling Water Return	**20.0** C	68 F
Differential Temperature	10.0 C	18 F
Surface Area for each Heat Exchanger	796 ft^2	
# of Heat Exchangers	2	
Flowrate of biofilter liquor through exchanger	119 gpm	
Flowrate of water through exchanger	119 gpm	
Turnover time in the biofilter	969 min	16.2 hrs

Energy Recovery and Generation

Total Heat Recovery from VerTad System	8,893,580 Btu/hr	
Heat Generation from VerTad Internal Recycle	0 Btu/hr	(At 122°F cooling water return)
Heat Removal from Biofilter	2,150,403 Btu/hr	
Heat used to Preheat Sludge to VerTad	6,743,178 Btu/hr	
Heat available from the Compressors	505,603 Btu/hr	

Table 4- : Capital Cost Estimate

Digestion

Category			Alternative WP1	Alternative WP2
Site Work			50,000	50,000
Demolition				
Anaerobic Digestion			-	-
Structures			1,390,000	930,000
Equipment & Mech			1,160,000	760,000
Patent Fee			-	-
Electrical/ I&C			340,000	270,000
Testing/ Start-up			incl	incl
NORAM provided equipment and engineering			7,010,000	4,130,000
Cased Reactors		1	6,640,000	3,660,000
Subtotal			16,590,000	9,800,000
Contractor Indirects, OH@P	35%		17,500	17,500
NORAM estimate	11.6%		1,105,480	651,920
NORAM provided	5%		350,500	206,500
Subtotal			18,063,480	10,675,920
Contingency	30%		1,441,642	943,159
Drilling	10%		722,975	398,713
NORAM provided	5%		381,630	224,957
Subtotal			20,609,727	12,242,749
Sales tax	8.4%		1,731,217	1,028,391
Subtotal			22,340,944	13,271,139
Allied Cost (35%)	35%		7,819,331	4,644,899
Less NORAM engineering	15%		(1,172,900)	(696,735)
Total			28,987,375	17,219,303